COMPARATIVE PERFORMANCE ANALYSIS OF DIFFERENT CONTROL STRUCTURE OF PROCESS

A THESIS

SUBMITTED IN PARTIAL FULFILMENT OF THE REQUIREMENT FOR THE DEGREE OF

MASTER OF TECHNOLOGY

IN

ELECTRONICS & INSTRUMENTATION

BY

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Rourkela-769 008, India

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UNDER THE GUIDANCE OF

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CERTIFICATE

This is to certify that the thesis entitled, "**Comparative Performance Analysis of Different Control Structures of Process**" submitted by **Mr. Rohit Khalkho** in partial fulfilment of the requirements for the award of Master of Technology Degree in Electronics and Communication Engineering with specialization in **"Electronics & Instrumentation"** during session 2011-13 at the National Institute of Technology, Rourkela (Deemed University) is an authentic work carried out by him under my supervision and guidance.

To the best of my knowledge, the matter embodied in the thesis has not been submitted to any other University/Institute for the award of any degree or diploma.

Prof. Tarun Kumar Dan

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Rohit Khalkho

ABSTRACT

This communication provides the comparative performance analysis of different control structures such as feedback only, feedback-feedforward, cascade and Internal Model Controller of process. The processes used in this project are Binary Distillation column, Heat Exchanger and Boiler Drum. Binary Distillation Column: Distillation column is a tall metal cylinder internally fitted with perforated horizontal plates used to promote separation of miscible liquids ascending in the cylinder as vapour. These columns function as process equipment where liquid or vapour mixture of two/more substance separation by application and removal of heat into its component fractions as per the desired purity requirement. Heat exchanger: It transfers the heat from a hot fluid to cooler fluid, so the primary importance of this system is to control the output temperature of heated liquid whenever there is some disturbances. Heat exchanger is generally used in chemical & process industries for efficient heating purpose of liquid steam where huge amount of heats are required & electrical heating is not economical. Boiler Drum: The primary purpose of the steam drum is to separate the saturated steam from the steam-water mixture that leaves the heat transfer surfaces and enters the drum. The steam-free water is re-circulated within the boiler with the incoming feed water for further steam generation. It is used to purify the steam by removing contaminants and residual moisture.

Comparative response analysis has been studied for different control structures (feedback only, feedback-feedforward and cascade). The responses are also compared using IMC based control system and effect of tuning parameters is also studied.

Key words: Feedback- Feedforward, cascade Internal Model Controller, Binary Distillation Column, Heat exchanger, Boiler Drum

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Chapter 1

INTRODUCTION

1.1 BACKGROUND

Distillation is a process in which a liquid or vapor mixture of two or more substances is substances is separated into its component fractions of desired purity by application and removal of heat [6]. It is well known that pure liquids exhibit different volatilities (i.e. vapor pressure) at a given temperature, and thus if heat is applied to a liquid mixture of these substances, the vapor so generated will be richer in the more volatile substances those having higher vapor pressures. If this vapor is condensed, it should be clear that a certain amount of purification will be achieved. This is the basic principle underlying a distillation operation [3].

A distillation process may be classified in one of two ways: Binary distillation refers to the separation of two substances and multi component distillation involves more than two substances.

The concept of relative gain array (RGA) to find correct pairings of controlled variables and manipulated variables and the concept of decoupling to achieve non-interacting feedback control.

A heat exchanger is such a equipment used for efficient way of heat transfer from one to another medium. The media that are used may be separated by a solid wall, so that they should never be mix, or they may not be in direct contact. They are mostly used in, refrigeration, space heating, natural gas processing, power plants, chemical plants, petrochemical plants, petroleum refineries, air conditioning and sewage treatment[4]. One of the fine examples of a heat exchanger is found in an IC engine in which a circulating fluid which is known as engine coolant passes through the coils of the radiator and air flows past the coils, due to this it cools the coolant and heat up the air. [4]

In the heat exchanger material, thinner the material between the fluid faster heats will transfer. Heat transfer is also affected by the surface area, thin material have the greater surface area means transfer more heat, in addition some fluids give up or accept heat more rapidly than other. Gases tend to transfer heat more than liquid because their molecules are further apart. The molecules take more time enriching each other to transfer heat, so molecules involves in heat transfer is very important.

Boiler drum- The primary purpose of the steam drum is to separate the saturated steam from the steam-water mixture that leaves the heat transfer surfaces and enters the drum. The steam-free water is re-circulated within the boiler with the incoming feed-water for further steam generation. The saturated steam is removed from the drum through a series of outlet nozzles, where the steam is used as is or flows to a super-heater for further heating.[5]

1.2 MOTIVATION

At the present time there are very crucial requirement of process control strategies with the satisfactory characteristics in the several process industries. Now a day the optimization of the process is very essential. Also the automation of the plant is must for the maintain the quality and brand of the product. These requirements come from the customer side because the market demand became more these days and we should know about the controllers characteristics for the different processes so it needs data sheet for this to provide the overview about the controllers.

1.3 LITRATURE REVIEW

For the different processes the various control strategies have been taken in this thesis. All the controllers that I have taken has its own advantage and disadvantages, we can find out the new control strategies using all these controllers .

Yuvraj, et al. [20] proposed the idea to control the temperature of the outlet fluid of conventional PID controller is developed.

Surekha, et al.[1] proposed the idea about the feedback, feed forward and cascade very well, also given the different control structures that can be implemented by own strategies.

B. Wayne, et al. [12] proposed the idea about the Boiler drum level control by using single-element, two-element and three-element control using simulink.

Chen, et al. [23] proposed a modified structure of internal model control for unstable process with time delay. They were design new structure using combination of feedback. feed forward, cascade and IMC control strategy.

Luyben, et al. [24] proposed the transfer function of distillation column. He was proposing the first and second order transfer function for distillation column. He was also developed an ATV method to determine the transfer function for nonlinear multivariable systems.

1.4 THESIS ORGANIZATION

Chapter-1 Introduction

Chapter-2 Processes

Chapter-3 Controllers

Chapter-4 Control Strategies

Chapter-5 Matlab Implementation and Results

Chapter-6 Conclusion & Scope for Future

Chapter 2
PROCESSES

2.1 DISTILLATION COLUMN
2.1.1 Introduction

Distillation is a process in which a liquid or vapor mixture of two or more substances is separated into its component fractions of desired purity by the application and removal of heat. Or Distillation is a method of separating mixtures based on differences in volatilities of components in a boiling liquid mixture. Distillation is a unit operation, or a physical separation process, and not a chemical reaction. Distillation is defined as a process in which a liquid or vapor mixture of two or more substances is separated into its component fractions of desired purity, by the application and removal of heat.**[6]**

The basic principle operation of a distillation column is that, it is known that liquids exhibit different volatilities (i.e., vapor pressure) at a given temperature, and thus if heat is applied to a liquid mixture of these substances, the vapor so generated will be richer in the more volatile substances those having higher vapor pressures.

2.1.2 Distillation equipment

The schematic diagram of a typical distillation column is shown below figure 2.1. The equipment consists of a vertical shell with a number of equally spaced trays mounted inside of it. Each tray contains two conduits, one on each side, called down comers. Liquid flows through these down comes by gravity from each tray to the one below. The vertical shell is connected by suitable piping to a heating device called a re-boiler. Reboiler to provide the necessary vaporization for the distillation process. The condenser to cool and condense the vapour leaving the column from top. The Reflux drum to held the condenser vapor from the top of column so that liquid(reflux) can be recycled back to the column. The vertical shell together with the condenser and reboiler constitute a distillation column.

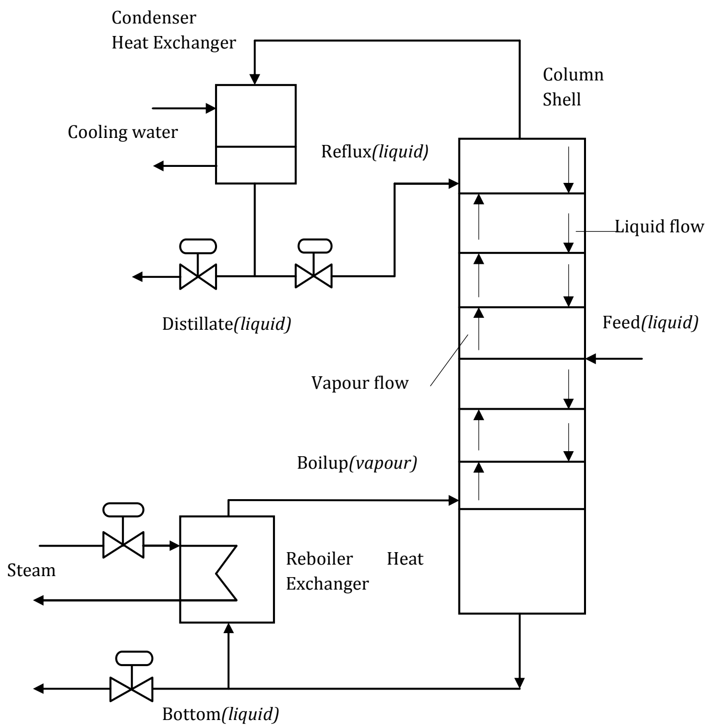

Figure 2.1 schematic of a typical distillation column[29]

2.1.3 Basic operation and terminology

The liquid mixture that is to be processed is known as the feed and this is introduced usually somewhere near the middle of the column to a tray known as the feed tray. The feed tray divides the column into a top (enriching or rectification) section shown in figure 2.2 and a bottom (stripping) section shown in figure 2.3. The feed flows down the column where it is collected at the bottom in the reboiler.

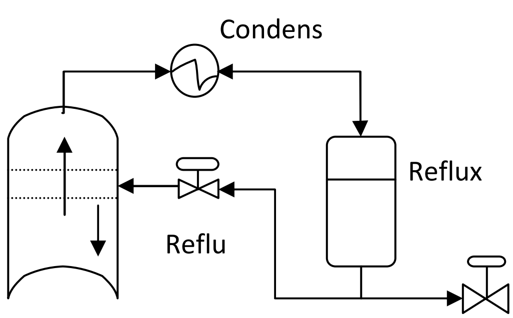

Figure 2.2 Enriching (or) rectification section[3]

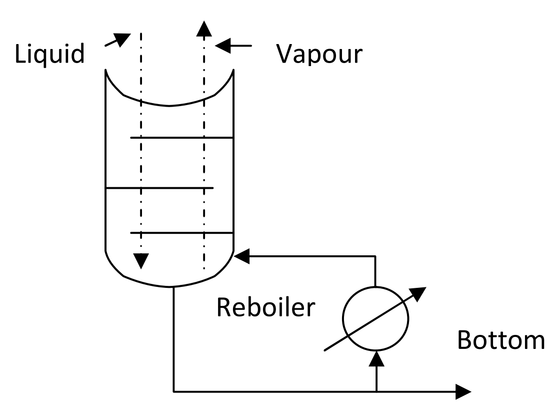

Figure 2.3 stripping section[3]

2.1.4 *Graphical Method For Binary Distillation*

Important design parameters in tray-type distillation column include the number of trays, binary distillation problems may be solved by either graphical or analytical methods. Graphical methods are McCabe-Thiele method and the Ponchon-Savarit method.

Separation of a binary mixture can be achieved in a single-stage process known as the equilibrium flash. If enhanced separation is desired, a column containing a suitable number of trays must be used.

Here we review only McCabe-Thiele method, which utilizes an x-y diagram.

2.1.4.1 *McCabe-Thiele method*

This method can be used when the following conditions are satisfied.

1. Molal heats of vaporization of the two substances are roughly the same
2. Heat effects (heats of solution, heat losses to and from column) are negligible.
3. These so-called constant-molal overflow assumptions imply that for every mol of vapor condensed, 1 mol of liquid is vaporized. Thus the liquid and vapor rates within each section of the tower remain constant.

The McCabe-Thiele method utilizes material balances and equilibrium relationships. These relationships are written for the enriching section and the stripping section and then combined to solve the binary distillation problem.

2.1.4.2 *Enriching and stripping section*

Referring to figure 2.4, a component material balance around a general tray *n* in the enriching section is written as:

$$V_n y_n = L_{n+1} x_{n+1} + D x_D \tag{2.1}$$

The trays are numbered from the bottom up. As per the assumption molal overflow, the subscripts of L and V are dropped. Equation (2.1) may be written as

$$V y_n = L x_{n+1} + D x_D \tag{2.2}$$

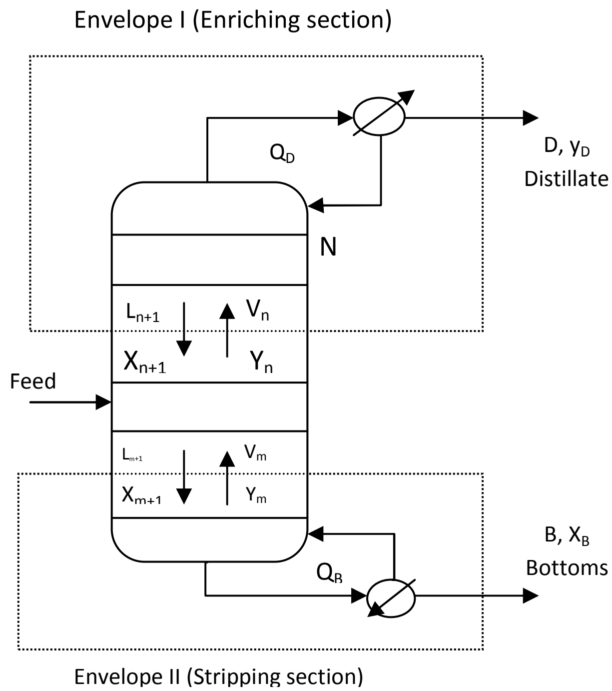

Figure 2.4 Material-balance envelopes for operating lines

Now solving for y_n to give

$$y_n = \frac{L}{V} x_{n+1} + \frac{D}{V} x_D \tag{2.3}$$

On x-y coordinates equation (2.3) is a straight –line of slope L/V with a y-intercept of x_D D/V. it relates the composition of the more volatile component in the vapour stream leaving a general tray *n* in the enriching section to that of the liquid entering tray *n*. This straight line, which represents the operating line for enriching section of the tower shown in Figure 2.5

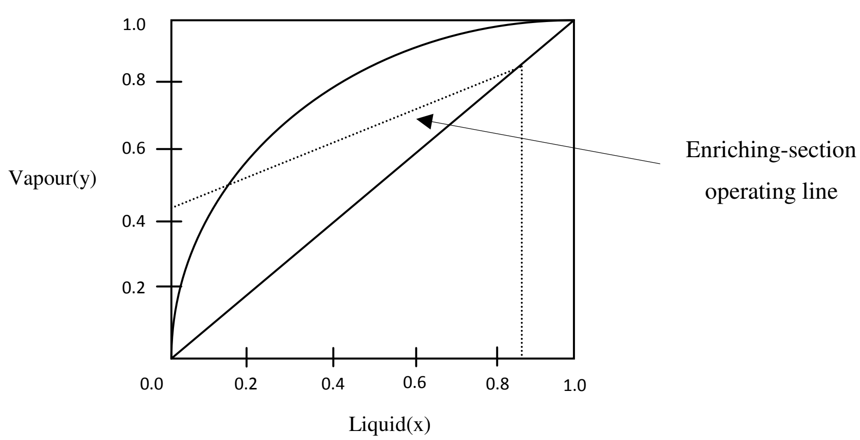

Figure 2.5 Enriching-section operating line

Similarly the equation for the stripping-section operating line can be developed by component material balance around envelope II in Figure 2.4

$$y_m = \frac{\bar{L}}{\bar{V}} x_{m+1} - \frac{B}{\bar{V}} x_B \tag{2.4}$$

Where the bar indicates that the stream is in the stripping section. This equation relates the composition of the more volatile component in the vapour stream leaving tray *m* in the stripping section to that in the liquid entering tray *m.* stripping section operating line has a slope of $\frac{\bar{L}}{\bar{V}}$ and it intersects the diagonal, where x=y, at x_B as shown in Figure 2.6.

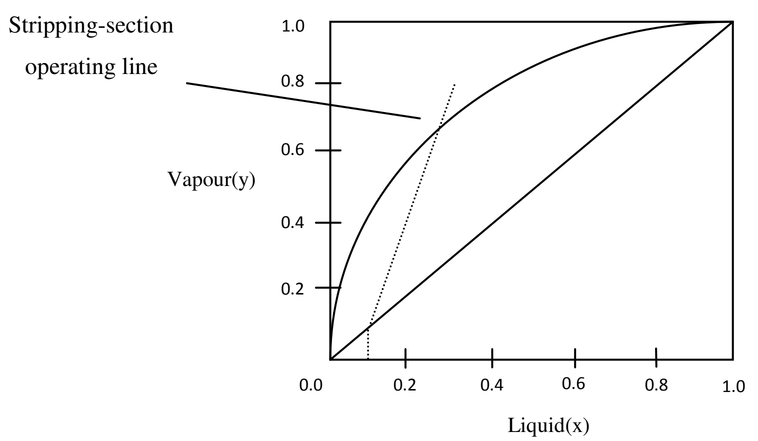

Figure 2.6 stripping-section operating line

2.1.4 DISTILLATION CONTROL CONCEPTS

Now we shift our focus to control concepts. Numerous questions arise in the development of control strategies for distillation towers. For example, how many variables should be controlled to achieve the production objectives? How should the manipulated and controlled variables be paired? Where should the sensors be located? Here we see the number of controlled variables in distillation.

In this section we present a method to determine the number of controlled variables in distillation.

Intuition suggests that the variables

- Product compositions
- Column pressure
- Base level
- Reflux accumulator level

Should be controlled for the following reasons:

1) Production objectives normally require the delivery of products of quality.
2) Product compositions to held at constant.
3) Column pressure is direct bearing on the separation capability of the tower.
4) An adequate inventory in the column base is desirable because without it the reboiler tubes would be exposed to excessively high temperature.
5) Finally a suitable liquid level in the reflux drum dampens out the effect of unexpected surges in vapor boilup upon reflux and serves as a liquid seal.

The model is based on unsteady-state mass balances. The assumption of constant molal overflow means that energy balances are not required we may write the following equations for the various section of the tower.

2.1.4.1 Condenser and reflux drum

Overall mass balance is

$$\frac{dM_D}{dt} = V - L_{N+1} - D \tag{2.5}$$

Where

M_D = molar holdup in reflux drum, mols

L_{N+1} = reflux, mol/time

D = distillate, mol/time

V = vapor flow, mol/time

Component mass balance is

$$\frac{dM_D x_D}{dt} = Vy_N - L_{N+1}\mathrm{x_D} - Dx_D \tag{2.6}$$

2.1.4.2 General Tray, n

Overall mass balance is

$$\frac{dM_n}{dt} = L_{n+1} - L_n \tag{2.7}$$

Where

M_n = liquid holdup on the n^{th} tray, moles

L_N = liquid leaving tray n, mol/time

Component mass balance is

$$\frac{dM_n x_n}{dt} = L_{n+1}x_{n+1} - L_n\mathrm{x_n} - Vy_{n-1} - Vy_n \tag{2.8}$$

2.1.4.3 Feed Tray ($n=N_F$)

Overall mass balance is

$$\frac{dM_{N_F}}{dt} = L_{N_F+1} - L_{N_F} + F \tag{2.9}$$

Where

M = liquid holdup on the feed tray, moles

L = liquid leaving feed tray, mol/time

Component mass balance is

$$\frac{dM_{N_F} x_{N_F}}{dt} = L_{N_F+1}x_{N_F+1} - L_{N_F}\mathrm{x}_{N_F} + Vy_{N_F-1} - Vy_{N_F} \tag{2.10}$$

2.1.4.4 Reboiler and column base

Overall mass balance is

$$\frac{dM_B}{dt} = L_1 - V - B \tag{2.11}$$

Where

M_B = molar holdup in column base, moles

B = bottoms flow, mol/time

L_1 = liquid leaving tray 1, mol/time

V = vapour boil up in recoiled, mol/time

Component mass balance is

$$\frac{dM_D x_D}{dt} = Vy_N - L_{N+1}\mathrm{x_D} - Dx_D \tag{2.12}$$

In addition to equation (2.5)-(2.12), each tray and the re boiler has an equilibrium relationship relating the compositions of affluent streams given by

$$y_n = \frac{\alpha x_n}{1+(\alpha-1)x_n} \tag{2.13}$$

2.2 HEAT EXCHANGER

2.2.1 Introduction

By adding or removing thermal energy Heat exchangers controlling a system's or substance's temperature. Although there are many sizes, sophistication levels, and types of heat exchangers, they all use a thermally conducting element usually in the form of a tube or plate to separate two fluids, such that one can transfer thermal energy to the other [7]. Home heating systems use a heat exchanger to transfer combustion - gas heat to water or air, which is circulated through the house. Power plants use locally available water or ambient air in quite large heat exchangers to condense steam from the turbines [7]. Many industrial applications use small heat exchangers to establish or maintain a required temperature. In industry, heat exchangers perform many tasks, ranging from cooling lasers to establishing a controlled sample temperature prior to chromatography. [7]

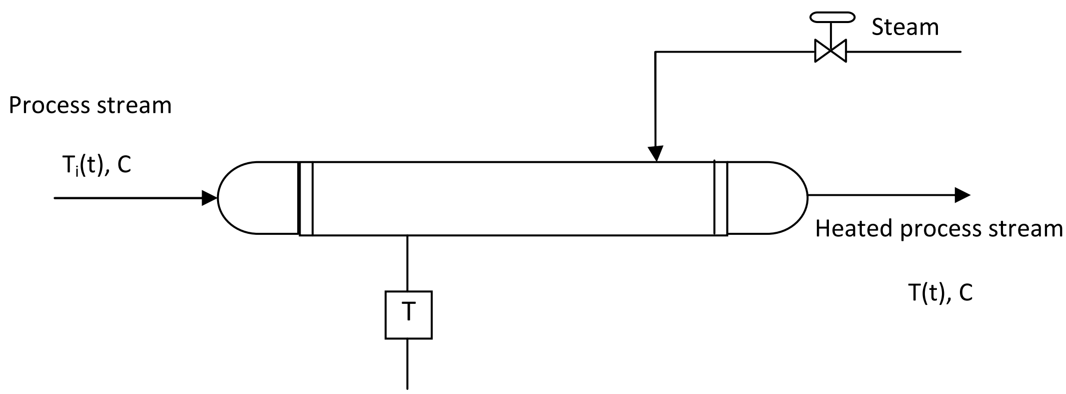

Figure: 2.7 Basic Heat Exchanger System

2.2.2 Fluid Fundamentals

How heat gets transferred from one fluid to another depends largely on the physical characteristics of the fluids involved, especially density, their specific heat, thermal conductivity, and dynamic viscosity. *Density* is a fluid's mass per unit volume, measured as lbm/ft3 (where lbm represents pounds of mass) or kg/m3. Density can be used to convert a measurement from a mass-flow rate, such as lbm/hr, to the more common volumetric units, such as gallons per minute for liquids, or cubic feet per minute for gases. Throughout a heat exchanger, the mass-flow rate remains constant, but changes in temperature and pressure can change the volumetric flow rate, particularly for a gas [7]. So a gas flow should be stated as a mass flow, a volumetric flow at standard conditions, or as a volumetric flow including temperature and pressure. In any case, the operating pressure should always be specified.

2.2.3 Fluid flow

Inside a heat exchanger, the fluid flow is either turbulent or laminar. Turbulent flow produces better heat transfer, because it mixes the fluid. Laminar-flow heat transfer relies entirely on the thermal conductivity of the fluid to transfer heat from inside a stream to a heat exchanger wall.[7]

An exchanger's fluid flow can be determined from its Reynolds number (N*Re*):

$$N_{\mathrm{Re}} = \frac{\rho \times v \times D}{\mu} \tag{2.14}$$

Where velocity and D flow is the diameter of the tube in which the fluid flows. The units cancel each other, making the Reynolds number dimensionless. If the Reynolds number is less than 2,000, the fluid flow will be laminar; if the Reynolds number is greater than 6,000, the fluid flow will be fully turbulent. The transition region between laminar and turbulent flow produces rapidly increasing thermal performance as the Reynolds number increases.[7]

The type of flow determines how much pressure a fluid loses as it moves through a heat exchanger. This is important because higher pressure drops require more pumping power. Although a manufacturer will normally determine the pressure drop, it is useful to predict the pressure drops that can occur with changing rates of flow. Laminar flow produces the smallest loss, which increases linearly with flow velocity. For example, doubling the flow velocity doubles the pressure loss. For Reynolds numbers beyond the laminar region, the pressure loss is a function of flow velocity raised to a power in the range 1.6–2.0. In other words, doubling the flow could increase the pressure loss by a factor of four.[7]

2.2.4 Heat Exchanger Equation

The heat-transfer rate (Q) of a given exchanger depends on its design and the properties of the two fluid streams. This characteristic can be defined as:

$$Q = UA\Delta T_{\log mean} \tag{2.15}$$

where U is the overall heat-transfer coefficient, or the ability to transfer heat between the fluid streams, A i s the heat-transfer area of the heat exchanger, or in other words the total area of the wall that separates the two fluids, and $\Delta T_{\log mean}$ is the average effective temperature difference between the two fluid streams over the length of the heat exchanger.[7]

A heat exchanger's performance is predicted by calculating the overall heat transfer coefficient *U* and the area *A*. The inlet temperatures of the two streams can be measured, which leaves three unknowns—the two exit temperatures and the heat- transfer rate[7]. These unknowns can be determined from three equations (the one above using an arithmetic average for $\Delta T_{\log mean}$ plus the heat-balance equation for each stream):

$$Q = UA\frac{(T_{inhot} - T_{outcold}) + (T_{outhot} - T_{incold})}{2} \tag{2.16}$$

$$= [\dot{m} \times Qc_p \times (T_{out} - T_{in})]_{cold} \tag{2.17}$$

$$= -[\dot{m} \times c_p \times (T_{out} - T_{in})]_{hot} \tag{2.18}$$

Solving these equations simultaneously usually requires iteration. In any case, a heat exchanger's manufacturer usually completes them.

2.3 BOILER DRUM

The boiler water/steam drum, or steam drum, is an integral part of the boiler's design. This vessel has three specific purposes; 8] provide a volume space to hold the boiling water in the boiler, 9] provide enough water volume to allow for good thermal mixing of the cooler bottom drum water with the hotter surface interface water, and 10] provide surface area and volume for the efficient release of the entrained steam bubbles from the boiler water. The surface area and volume of the vapor space in the water/steam drum is critical to the efficient separation of the steam bubbles from the water. Too small an area can result in an excessive surface tension and high velocities, which result in wasted heat and drum water carry-over. Too large an area is simply a waste of materials and labor to construct the vessel. The boiler water/steam drum also provides a logical location for 8] addition of feedwater,9] addition of chemical water treatment and 10] surface blow down, which helps reduce the surface tension of the water/steam interface to allow better steam release. Because all of these tasks involve the removal and addition of some mass (water or steam) the water/steam interface is always in a state of flux. This lack of circulation also reduces the effectiveness of the chemical water treatment and can cause precipitation of the chemicals as chemical salts or foams. High water levels raise steam exit velocities and result in priming or boiler water carryover in to the distribution system. Priming results in wet dirty steam while carry-over can result in dangerous water hammer and pipe or equipment damage.[5]

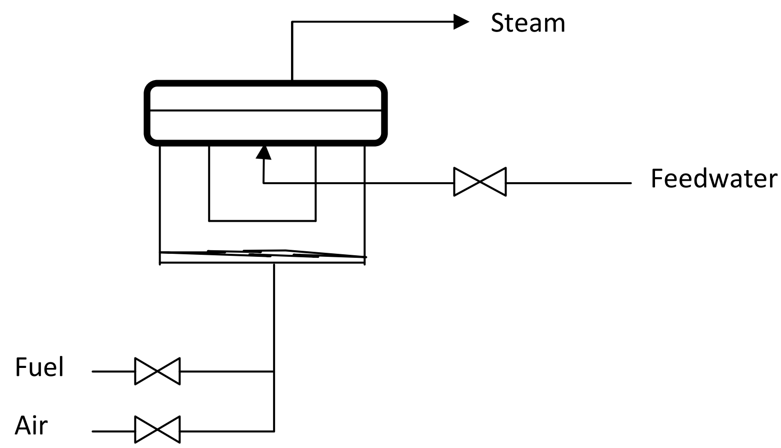

Figure: 2.8 Schematic of Boiler Drum

2.3.1 Liquid level measurement

Measurement of liquid level provides us to maintain the liquid level with the reference water level of the boiler drum. It is very necessary to maintain the level of the liquid for the better operation.

2.3.1.1 Inferential Level Measurement

In the inferential level measurement the method is that pressure (static) applied by any liquid is directly proportional to the height of the liquid above the point of measurement, irrespective of volume [5]. Thus, if there is some pressure difference occurs in any instrument then we can easily calibrate in terms of liquid height.

2.3.1.2 Volumetric Level Measurement

In the Volumetric Level Measurement the level of the liquid can be calibrated by using the float in the vessel which transmits the liquid level against the surface of the liquid. It is the direct method finding the liquid level by the means of up and down of the float on the liquid. In this method, float can be used to transform the displacement into electrical signal that can be analyzed easily.

2.3.1.3 Differential Pressure Level Measurement

In the differential pressure measurement level of the liquid can be measured by the use of differential pressure device whose range should be equal to head of the liquid to be measured. By this method Liquid levels in closed tanks may be measured under pressure.

Chapter 3

CONTROLLERS

3.1 FEEDBACK CONTROLLER

In feedback control system when output signal is deflected by the external or internal disturbance then this signal is compared with the reference signal and generated error signal is fed to the controller which takes the corrective action to give the proper output.

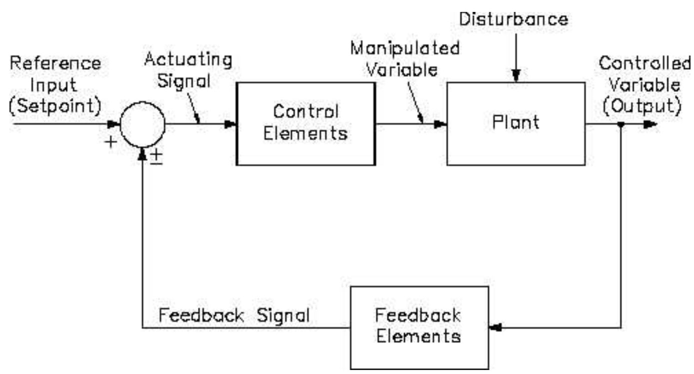

Figure: 3.1 Block Diagram of feedback control

3.2 FEEDFORWARD CONTROLLER

Conventional feedback control loops can never achieve perfect control. It is difficult for the conventional loops to keep the process output continuously at the desired set point value in the presence of load or set point changes. This is because a feedback reacts only after it has detected a deviation in the value of the output from the desired set point. Unlike the feedback control systems, a feed forward control configuration measures the disturbance directly and takes the control action to eliminate its impact on the process output.

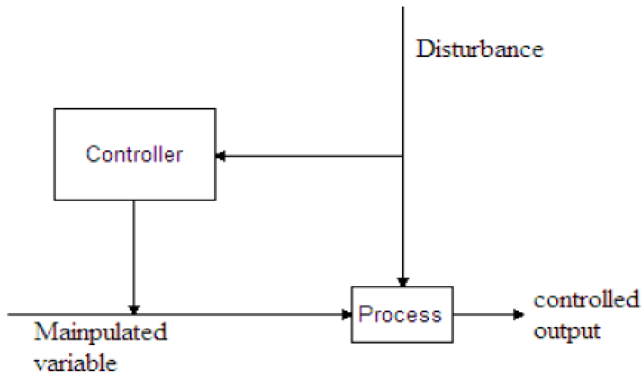

Figure : 3.2 Basic block diagram of feedforward control

In feedforward control strategy, information concerning one or more conditions that might disturb the control variable is converted into corrective action to minimize deviation of controlled variable. The signals which have the potential to upset the process are transmitted to the controller. The controller makes appropriate computation on these signals and calculates new values for the manipulated signals and sends those to the final control element, Therefore, the control variable remains unaffected in spite of load changes.

3.3 CASCADE CONTROLLER

In cascade control configuration, we have one manipulated variable and more than one measurement. Cascade control uses the output of primary controller to manipulate the set point of secondary controller.

The basic principle of cascade control is that if the secondary variable responds to the disturbance sooner than the primary variable, then there is a possibility to capture and nullify the effect of the disturbance before it propagates into the primary variable. The concept of cascade control is shown below Figure(3.3).

The two measurements are taken from the system and used in their respective control loops. In the outer loop, the controller output is the set point of the inner loop. The outer loop is called primary loop and the inner loop is called secondary loop. Thus, if the outer loop dynamic variable changes, the error signal (i.e., input to the controller) affects a change in set point of the inner loop. Even though the measured value of the inner loop has not changed, the inner loop experiences an error signal and produces a new output by virtue of the set point change. The primary objective of cascade control is to divide an difficult control process into two portions: whereby a secondary control loop is formed around major disturbances leaving only minor disturbances to be controlled by the primary controller.

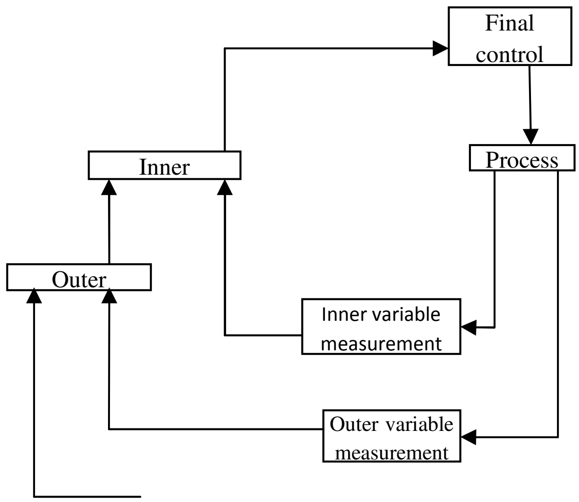

Figure: 3.3 Cascade control system

3.3.1 Advantage of Cascade control

Cascade control is most advantageous in applications where the secondary closed loop can include the major disturbance, and only the major lag is included in the primary loop. Some advantages of cascade control are[1]

1. Better control of the primary variable
2. Primary variable less affected by disturbances
3. Faster recovery from disturbances
4. Increase the natural frequency of the system
5. Reduce the effective magnitude of a time-lag
6. Improve dynamic performance
7. Provide limits on the secondary variable

3.4 INTERNAL MODEL CONTROLLER (IMC)

In industrial system, internal model control has proved to be successful controller design strategies. It provides a diaphanous variant for the tuning and designing of various control architecture .It is a commonly used technique. Cumulatively it gives good set point tracking and set point tracking with a time-constant ratio. But disturbance rejection has a greater priority than set point tracking in various control applications. Hence the controller design that emphasizes disturbance rejection rather than the set point tracking is an important criterion that must be taken into consideration.[21] Internal model control is an advance control technique in which process model is used in order to compute the value of control variable[21]. In internal model control process model is connected in parallel with the actual process, with the help of this we compare both of process.[21]

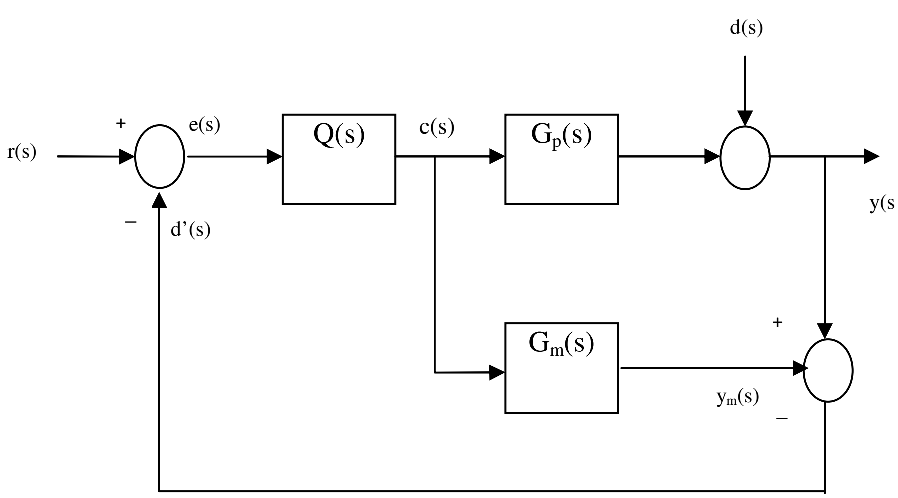

Figure: 3.4 Internal Model Control structure

Where:
$Q(s)$ = primary controller (IMC) T/F.
$G_p(s)$ = process T/F.
$G_m(s)$ = process model T/F.
$r(s)$ = set point
$e(s)$= error.
$c(s)$= manipulated variable.
$d(s)$ = disturbance.
$y_m(s)$ = model output.
$y(s)$ = controlled variable (process output).

Chapter 4
Control Strategies

4.1 DISTILLATION COLUMN CONTROL STRATEGIES

Here we will see how and why certain variables may be manipulated to control one or both product compositions in a distillation tower. For simplicity we will consider the separation of a single binary feed F into two products D and B. schematic of the tower is shown in the Figure 3.1.

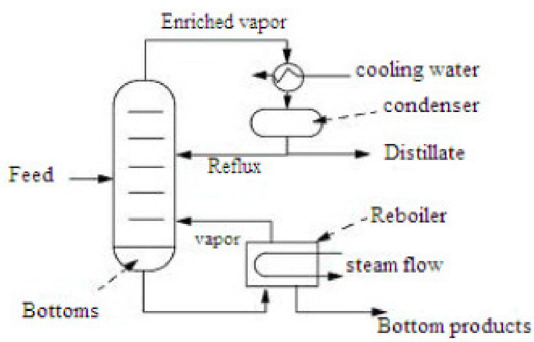

Figure 4.1 schematic of a distillation tower

The overall material balance for this column may be expressed as

$$F = D + B \tag{4.1}$$

The component material balance for the more volatile substance, which we donate as substance A, is

$$Fx_F = Dx_D + Bx_B \tag{4.2}$$

The term B in the equation (4.2) may be replaced by F-D in accordance with equation (4.1) to give

$$Fx_F = Dx_D + (F - D)x_B$$

Or

$$\frac{D}{F} = \frac{x_F - x_B}{x_D - x_B} \tag{4.3}$$

Equation (4.3) gives the unique steady-state relationship between D/F and x_D, x_B, x_F.

Similarly, if we had replaced D in the equation (4.2) by F-B, we will get

$$\frac{B}{F} = \frac{x_D - x_F}{x_D - x_B} \tag{4.4}$$

Equation (4.4) gives the unique steady-state relationship between B/F and x_D, x_B, x_F.

Typically the control objectives in a distillation operation are to maintain x_D and/or x_B at set point in the presence of disturbances. These disturbances may be characterized as (1) process loads, (2) changes in cooling- and heating-medium supply conditions, and (3) equipment fouling.

Our control objectives are to maintain x_D and/or x_B constant in the presence of changes in F and x_F.

Consider the separation of a binary feed F having the composition x_F into two products D and B that have compositions x_D and x_B, respectively, in a distillation column with N ideal stages. Assume that the feed is a saturated liquid and that the relative volatility of the binary system is constant. If we further assume, that the concept of constant molal overflow applies, then we know that the required separation can be achieved in N ideal stages s shown in Figure 4.2.

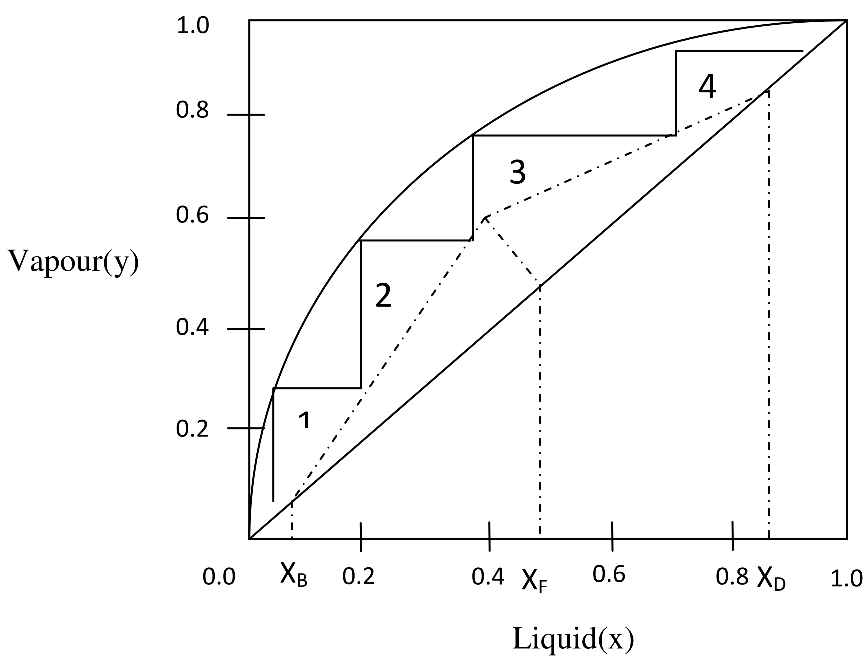

Figure 4.2 McCabe-Thiele diagram

Thus if a saturated liquid feed F_s with composition x_{Fs} is feed to a distillation column having N ideal stages operating at steady state and the reflux rate is L_s and the vapour boil up V_s, then the column will produce two products, D_s and B_s, that have compositions x_{Ds} and x_{Bs} respectively.

But distillation columns are not always at steady-state because of the disturbances. Therefore, we need to know which variables to manipulate to control x_D and/or x_B at their steady state values in the presence of disturbances.

To solve this control problem, we begin with the examination of equation (4.3):

$$\frac{D}{F} = \frac{x_F - x_B}{x_D - x_B} \tag{4.3}$$

This equation suggests that for a given F and x_F, a change in D affects x_D and/or x_B. Therefore, we may postulate that when changes in F or x_F occur, it may be possible to manipulate D so as to control x_D or x_B. But here in equation (4.3) there are two unknowns (x_D and x_B) so we cannot assess the quantitative effects of changes of D upon x_D and/or x_B.

Another relationship is found from McCabe-Thiele diagram between vapour or (reflux) flow to the two compositions. Several investigators derived relationship from many examines.

One such relationship is

$$\frac{V}{F} = \frac{(R_m+1)\frac{D}{F}}{1-1.6612\left[\frac{\frac{\ln S}{\ln \alpha}+1}{N+1}-0.25\right]^{1.7643}} \tag{4.5}$$

Where R_m is the minimum reflux ratio given by

$$R_m = \left(\frac{1}{\alpha-1}\right)\left(\frac{x_D}{x_F} - \frac{\alpha(1-x_D)}{1-x_F}\right) \tag{4.5a}$$

And S is called "separation factor," which is defined as

$$S = \frac{x_D(1-x_B)}{x_B(1-x_D)} \tag{4.5b}$$

Equations (4.3) and (4.5) completely describe the effect of changing D and/or V (or equivalently D/F or V/F) upon x_D and x_B. For a given upset in F or x_F, these equations will tell us how to control x_D and/or x_B.

4.1.1 Pairing and Interaction in Distillation

Here we see a systematic procedure for determining the correct pairing of controlled and manipulated variables. As we note that with the column on pressure control, there are four controlled variables-x_D, x_B, h_D, and h_B and four manipulated variables- D, B, L_{N+1}, and V. Thus there are 4! Or 24 possible ways of pairing these variables. In dual composition control, all four controlled variables are connected to the four manipulated variables.

For dual composition control: We assume that the column pressure is constant. Then the column has two inventory control loops and two composition control loops. The composition control loops can interact among themselves, and thus this is a multivariable control problem.

Several approaches for the solution of multivariable control problems are available. The best known is based on the concept of relative gain to find correct pairings of the controlled and manipulated variables and the concept of decoupling to achieve non-interacting feedback control.

We consider the problem of control system design for dual composition control in two parts. We first determine the proper pairings of controlled and manipulated variables in dual composition control, and then focus on the problem of interaction.

Consider a general 2x2 process:

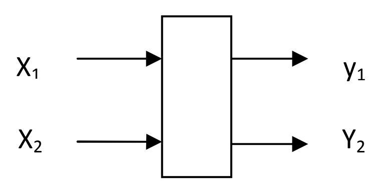

Figure 4.3 A general 2x2 process

The block diagram of a general 2x2 system is shown in Figure 4.3 the controlled variables are Y_1, Y_2 and the manipulated variables are X_1, X_2[1].

The relationships among these variables around some steady-state operating point may be expressed as:

$$\Delta Y_1 = K_{11}\Delta X_1 + K_{12}\Delta X_2 \tag{4.6}$$

And

$$\Delta Y_2 = K_{21}\Delta X_1 + K_{22}\Delta X_2 \tag{4.7}$$

The K's are steady-state gains that can be determined from mathematical models or from experimental tests. They describe how, say, X_1 affects Y_1 when Y_2 is not controlled.

A second gain may be defined that gives a measure of how, say, X_1 would affect Y_1 if Y_2 were under close-loop control by the relationship:

$$a_{11} = \left.\frac{\Delta Y_1}{\Delta X_1}\right|_{Y_2\ constant} \tag{4.8}$$

The gain a_{11} can be determined by setting $\Delta Y_2 = 0$ in the equation (4.7), solving for ΔX_2 and substituting it in the equation (4.6), and solving for $\frac{\Delta Y_1}{\Delta X_2}$. The other gains a_{12}, a_{21}, a_{22} can be determined in a similar manner. The ratio of K_{ij} to a_{ij} is called the relative gain λ_{ij}. Thus,

$$\lambda_{ij} = \frac{K_{ij}}{a_{ij}} \tag{4.9}$$

Where i=1,2 and j=1,2

$$\lambda_{ij} = \frac{K_{ij}}{a_{ij}} = K_{ij}C_{ij} \tag{4.10}$$

Once the relative gains are calculated, they are arranged in a matrix form:

$$\lambda = \begin{bmatrix} \lambda_{11} & \lambda_{12} \\ \lambda_{21} & \lambda_{22} \end{bmatrix} \tag{4.11}$$

Comparison of the relative gains for each manipulated variable gives us quantitative information on which input X has the most influence on a given output Y, and consequently how to pair the X's and Y's.

For example Consider the wood-Berry model [3]

$$\begin{bmatrix} X_D(s) \\ X_B(s) \end{bmatrix} = \begin{bmatrix} \frac{12.8e^{-s}}{16.7s+1} & \frac{-18.9e^{-3s}}{21s+1} \\ \frac{6.6e^{-7s}}{10.9s+1} & \frac{-19.4e^{-s}}{14.4s+1} \end{bmatrix} \begin{bmatrix} R(s) \\ S(s) \end{bmatrix} + \begin{bmatrix} \frac{3.8e^{-8.1s}}{14.9s+1} \\ \frac{4.9e^{-3.4s}}{13.2s+1} \end{bmatrix} F(s)$$

The process gain matrix K is

$$K = \begin{bmatrix} 12.8 & -18.9 \\ 6.6 & -19.4 \end{bmatrix}$$

And

$$C = (K^{-1})^T = \begin{bmatrix} 0.157 & -0.153 \\ 0.053 & -0.104 \end{bmatrix}$$

Now RGA (λ) is K_{ij} multiplied with C_{ij} (element to element multiplication)

Relative Gain Array (λ) = $\begin{bmatrix} 2.0094 & -1.0094 \\ -1.0094 & 2.0094 \end{bmatrix}$

For each controlled variable, the manipulated variable selected is the one with the largest positive relative gain, closest to 1.

i. Relative gains numerically close to one another indicate a highly interacting system.
ii. If an X and a Y having negative relative gain are paired, the system will be uncontrollable and unstable.

So here for the above example of Wood –Berry model the relative gains that x_D-R, and x_B-S are workable conditions. Thus the Distillation composition is controlled by manipulating the reflux flow R and the bottoms composition is controlled by the manipulating steam flow S.

4.1.2 Cascade Control Distillation Column

In distillation operations the disturbances occur in the condenser and in the re boiler. The condenser coolant and/or the re boiler heating medium are often process fluids in other parts of the plant and any disturbance on the process side in that portion of the plant can manifest themselves as disturbances associated with the condenser and/or the re boiler.

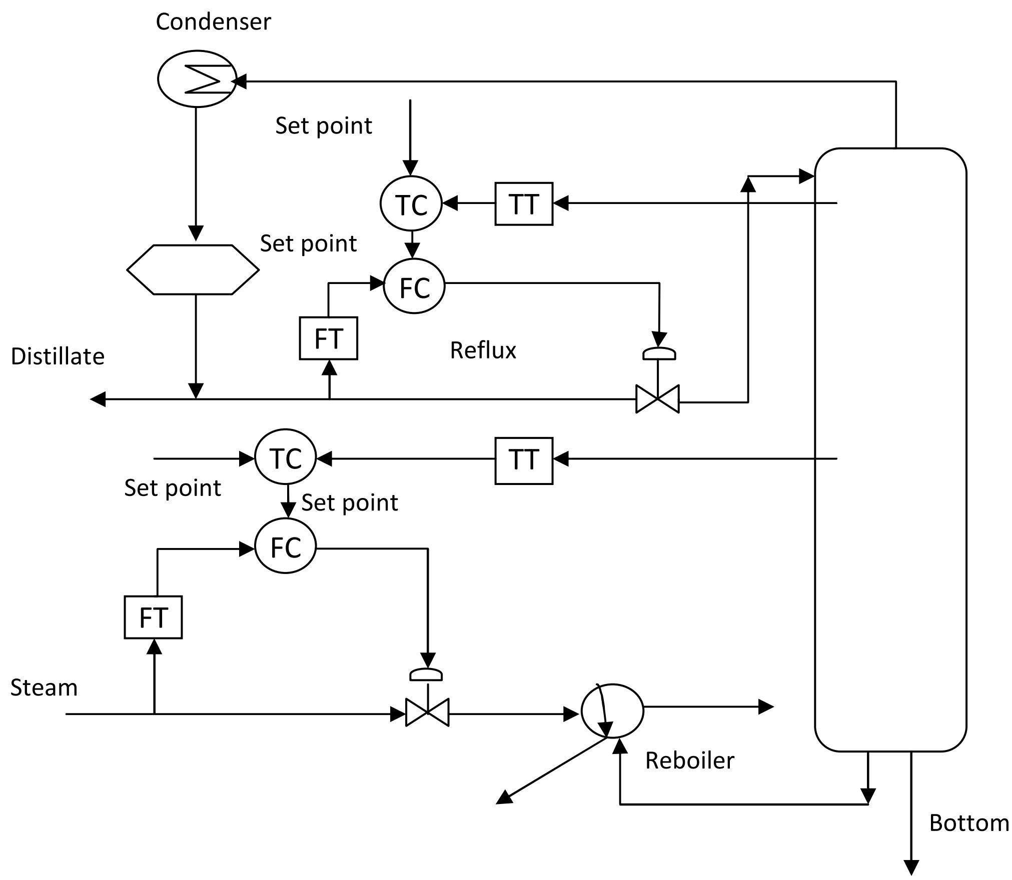

Figure 4.4 cascade control of distillation column

Let us see how these disturbances affect controlled variables. Figure 4.3 shows a schematic of the distillation column control system it is assumed that the bottoms composition is inferentially controlled by controlling a suitable tray temperature at set point. For this loop steam flow is the manipulated variable. If the steam supply pressure fluctuates, the pressure will drop across the control valve, thus the flow of the steam to the reboiler fluctuates. These fluctuations upset the control-tray temperature. When the temperature controller senses the upset, it takes the correlative action. The purpose of the second loop is to adjust the manipulated variable whenever disturbance in that loop takes place so that the performance of the outer loop remains unaffected. Thus, for the example shown in Figure 4.4, if the steam supply pressure changes, the pressure drop across the orifice changes. The orifice transmitter senses the change and the flow controller manipulates the valve position so as to maintain the flow of steam constant.

Control systems such as these in which the output of one controller serves as the set point of another controller are called cascade control systems. The inner loop consisting of the orifice sensor/transmitter, the flow controller, and the control valve is referred to as the secondary loop.

The outer loop consisting of the temperature sensor/transmitter and temperature controller along with the elements of inner loop is referred to as the primary loop.

Success with cascade control is possible only if the inner loop is at least as fast as, and preferably much faster than, the outer loop. Outer loop must compensate for the disturbances in that loop rapidly enough that the outer loop is unaware that they are occurring.

4.1.3 Feedforward Control of Distillation Column

Figure shows the Distillation column with feed forward control. The product comes out from the top of the distillation column is called top product and The product comes out from the bottom of the distillation column is called bottom product. Hot stream comes from top is cooled by condenser, while re-boiler heats up the bottom product. There are two disturbances the composition, C_f, and the feed flow rate, F_f. The reflux ratio and the re-boiler stream pressure are the two existing manipulated variables. Reflux ratio depicts that how much amount of the top product to be fed back to distillation's top. The available control variable is bottom product.

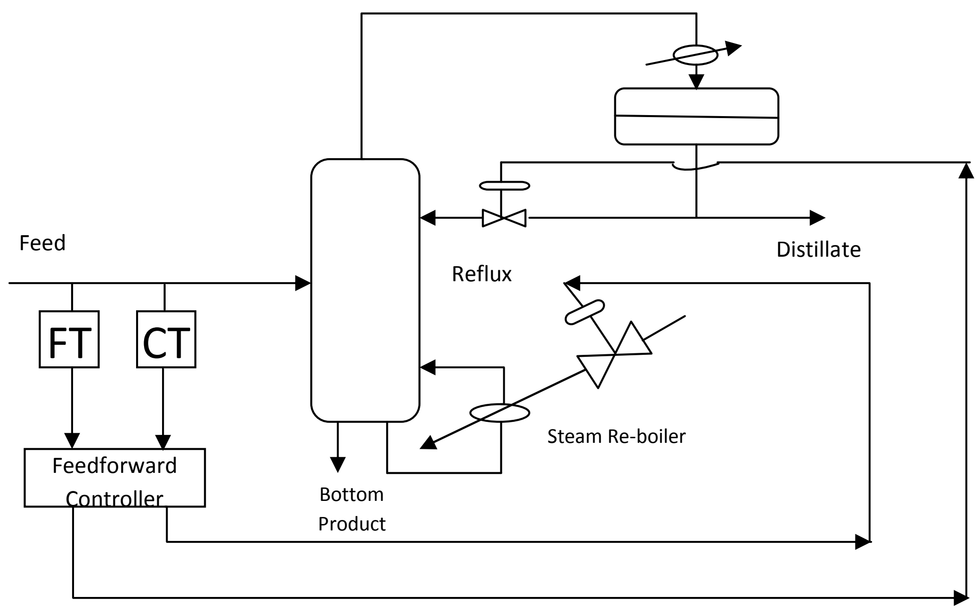

Figure: 4.5 Feed forward control of distillation colum

4.2 HEAT EXCHANGER CONTROL STRATEGIES

For the sake of convenience there are different assumption have been taken in this paper. The outflow and inflow rate of fluid will be same is the first assumption, so that the level of the fluid in the heat exchanger is maintained. Second is the heat absorbing capacity of the separating material or wall is negligible. As sensing element a thermocouple is used, which is introduced in the feedback path of the control scheme. Temperature measured by the thermocouple supplied to the transmitter unit to convert the voltage in the range of 4-20 mA to a standardized signal. This standardized signal is used by the controller unit. Then the controller applies the control strategy and compares set point with the output signal and then gives required command signal through the actuator to the final control element.[11]

Transfer function given below have taken from the paper.[11]

These transfer functions have been used for the different control strategies such as only Feedback, Feedback-feed forward, Cascade and Internal Model Controller. All controllers have realised by its basic steps of controller formation. After applying basic steps on the given structures of controller it gives a final equation and response of this final equation have obtained by the MATLAB implementation.

Transfer function of process (G_{p1}) $\dfrac{50e^{-sT_d}}{30s+1}$ (4.12)

Gain of valve 0.13 (4.13)

Transfer function of valve (G_{p2}) $\dfrac{0.13}{3s+1}$ (4.14)

Gain of I/P converter 0.75 (4.15)

Transfer function of disturbances (G_{d1} , G_{d2}) $\dfrac{1}{30s+1}$, $\dfrac{3}{30s+1}$ (4.16)

Transfer function of thermocouple (feedback components) (G_{m1}) $\frac{0.16}{10s+1}$ (4.17)

4.2.1 Feedback control of heat exchanger

The feedback control measures the temperature of the variable directly which is to be controlled. The controller decreases the amount of inlet steam If the outlet water temperature is so high. Likewise, the controller increases the amount of inlet steam If the outlet water temperature is too low. This is the example of negative feedback to maintain the outlet temperature at the set point.

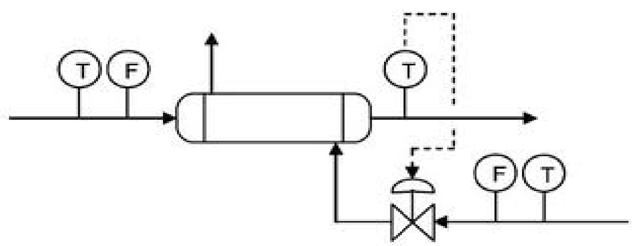

Figure: 4.6 Schematic of feedback control strategies of heat exchanger

4.2.2 Feedback-feed forward control of heat exchanger

In the combination both controller controlled variable is the outlet temperature and main load variable - the inlet temperature. Set point to the steam pressure controller is combined output of the both controller.

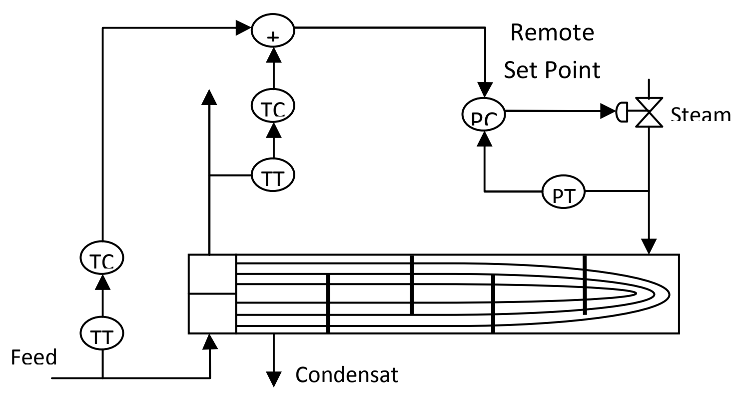

Figure: 4.7 schematic of combined feedback-feed forward controller

Figure shows the generalized form of feedback-feed forward control.

Let's develop an equation of the composite system for the closed loop response.

$$\bar{y} = G_p \bar{m} + G_d \bar{d}$$

Manipulated variable is given by:

$$\bar{m} = G_f \bar{c} = G_f G_{c1} \bar{e}_1 + G_f G_{c2} \bar{e}_2 \tag{4.18}$$

$$\bar{m} = G_f G_{c1}(\bar{y}_{sp} - G_{m1}\bar{y}) + G_f G_{c2}(G_{sp} y_{sp} - G_{m2}\bar{d}) \tag{4.19}$$

We got using above equation:

$$\bar{y} = \frac{G_p G_f (G_{c1} + G_{c2} G_{sp})}{1 + G_p G_f G_{c1} G_{m1}} \bar{y}_{sp} + \frac{G_d - G_p G_f G_{c2} G_{m2}}{1 + G_p G_f G_{c1} G_{m1}} \bar{d} \tag{4.20}$$

From above equation we can realise the following related to closed loop process response:

1. Stability of the closed loop using by finding the roots of the characteristics equation:

$$1 + G G_f G_{c1} G_{m1} = 0 \tag{4.21}$$

2. Transfer function of feed forward loop:

$$G_{c2} = \frac{G_d}{G_p G_f G_{m2}} \qquad and \qquad Gsp = \frac{G_{m2}}{G_d} \tag{4.22}$$

3. $Gd - Gp\ Gf\ Gc2\ Gm2 \neq 0 \quad and \quad Gp\ Gf\ Gc2\ Gsp \neq 1$ (4.23)

If any of these parameters (G_d G_p, G_f, G_{m2}) is approximately known.[1]

4.2.3 Cascade control of heat exchanger

The control variable controls the opening of the hot water valve. The primary loop controls the product temperature. The secondary loop controls the heat flow to compensate for flow variations (disturbances). The valve with flow control system can be regarded a new valve with an approximate proportional relation between the control variable and the heat flow.

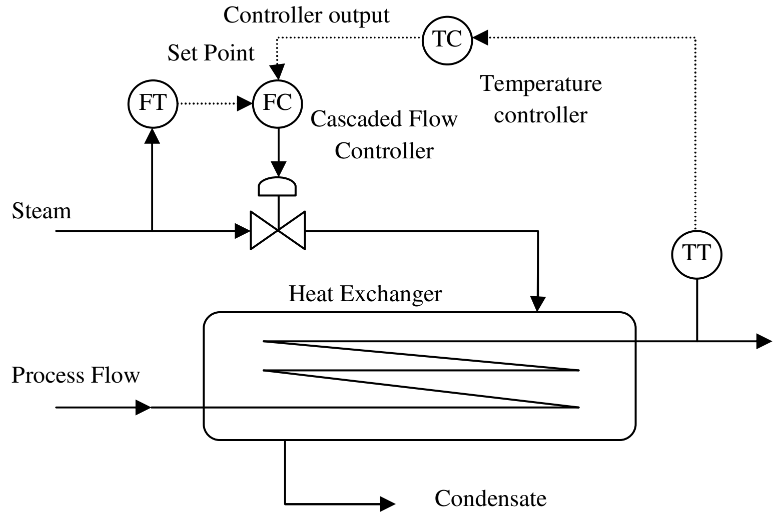

Figure: 4.8 Schematic of the cascade control of heat exchanger

G_{c1} – Primary controller

G_{c2} – Secondary controller

G_{p1} – Primary process

G_{p2} – Secondary process

G_{m1} and G_{m2} – Measuring device

G_{l1} and G_{l2} – Disturbance gains

L_1 and L_2 – Disturbance inputs

R1 – Set point

C_1 – Primary controlled variable

C_2 – Secondary variable

Mason's gain formula $\frac{y}{x} = \frac{1}{\Delta}\sum_{k=1}^{N} P_k \Delta_k$ [1] (4.24)

Where Y is the output variable, and X is the input variable.

N = total number of forward paths

P_k= path gain of k^{th} forward path

Δ = 1- (sum of loop gains of all individual loops) + (sum of gain products of all possible combinations of two non-touching loops) - (sum of gain products of all possible combinations of three non-touching loops) + …

Δ_k = the value of Δ for that part of the graph not touching the k^{th} forward path [1]

$$\frac{C_2}{R_2} = \frac{G_{c2}(s)G_{p2}(s)}{1+G_{c2}(s)G_{p2}(s)G_{m2}(s)} \tag{4.25}$$

$$\left(\frac{C_1}{R_1}\right) = \frac{G_{c1}(s)G_{c2}(s)G_{p1}(s)G_{p2}(s)}{1+G_{p2}(s)G_{m2}(s)G_{c2}(s)+G_{p1}(s)G_{p2}(s)G_{c1}(s)G_{c2}(s)G_{m1}(s)} \tag{4.26}$$

Cascade

$$\left(\frac{C_1}{R_1}\right) = \frac{G_{c1}(s)G_{c2}(s)G_{p1}(s)G_{p2}(s)}{1+G_{p1}(s)G_{p2}(s)G_{c1}(s)G_{m1}(s)}; G_{c2} = 1, G_{m2} = 0 \tag{4.27}$$

Simple feedback

Assuming that major disturbances enter the secondary loop:

$$\left(\frac{e}{L_2}\right) = \frac{-G_{l1}(s)G_{p1}(s)G_{m2}(s)}{1+G_{p2}(s)G_{m2}(s)G_{c2}(s)+G_{p1}(s)G_{p2}(s)G_{c1}(s)G_{c2}(s)G_{m1}(s)} \tag{4.28}$$

Cascade

$$\left(\frac{e}{L_2}\right) = \frac{-G_{l2}(s)G_{p1}(s)G_{m1}(s)}{1+G_{p1}(s)G_{p2}(s)G_{c1}(s)G_{m1}(s)} \tag{4.29}$$

Conventional feedback

These expressions can be used to analyse the effect of adding a secondary loop to the basic feedback loop in terms of performance, such as state error and speed of response.[1]

4.2.4 Internal model control for heat exchanger

The figure shows the IMC structure. In this model process model distinguishes the characteristics and is parallel with the actual process. Where bar represents the process model.

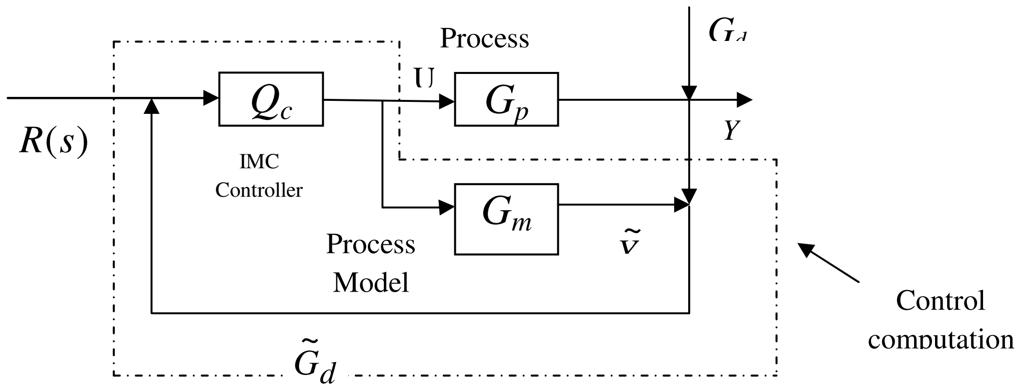

Figure: 4.9 Block diagram of IMC control strategies[12]

Signal to the controller is:

$$\tilde{R} = R - \big(G_P(s) - G_m(s)\big)U(s) - G_d \tag{4.30}$$

“Internal model control works with different condition which we given below.

Perfect Model, Without Disturbances: $G_p(s) = G_m(s)$ & $d(s) = 0$ (4.31)

$$y(s) = G_p(s) \times Q(s) \times R(s) \tag{4.32}$$

Perfect Model , With Disturbance:

$$y(s) = [G_P(s)Q(s)R(s) + \{1 - G_m(s)Q(s)\}d(s)] \tag{4.33}$$

With only Disturbance & Disturbance Rejection:

$$y(s = [1 - G_m(s)Q(s)]d(s) \quad [12] \tag{4.34}$$

4.3 Boiler Level Control Strategies

There are three types of control strategies for the Boiler Drum level single-element , two-element and three-element control. These all are based on the different size of load of the Boiler Drum.

4.3.1 Single-element control

In the single element control it uses the single measuring device which facilitates the control signal to the regulator of the feed-water. This control strategy is based on the ON/OFF control of the feed-water.

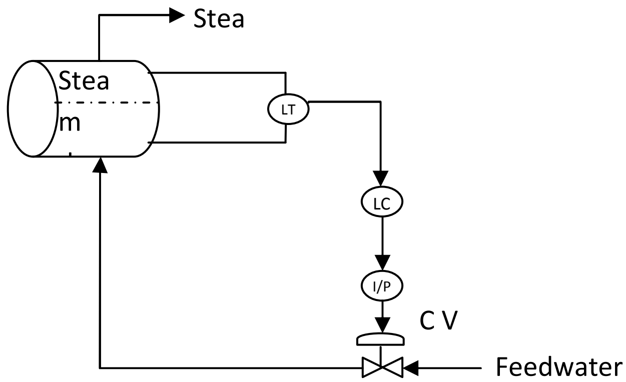

Figure: 4.10 schematic of single-element control

4.3.2 Two–element control

The two-element control can be used for the any size of the boiler Drum. In this strategy there are two variables, drum level and feed-water supply.

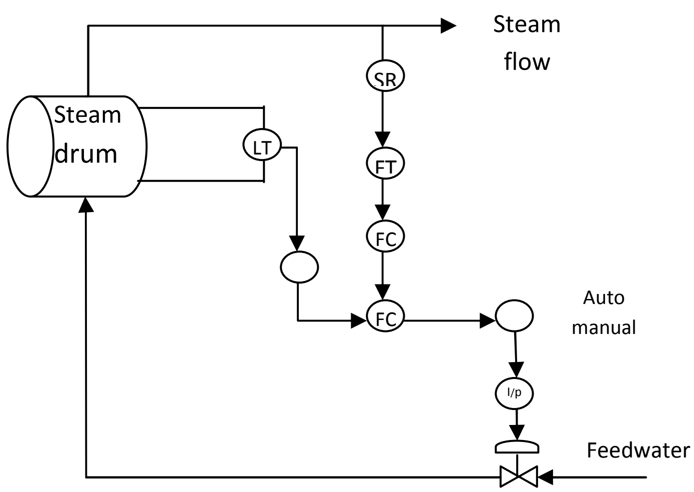

Figure: 4.11 schematic of two-element control

4.3.3 Three-element control

The two-element control still needs some modification so in this one more variable is added that is the flow of the feed water. The output of the two-element is cascaded with the third manipulated variable that is flow of feed-water.

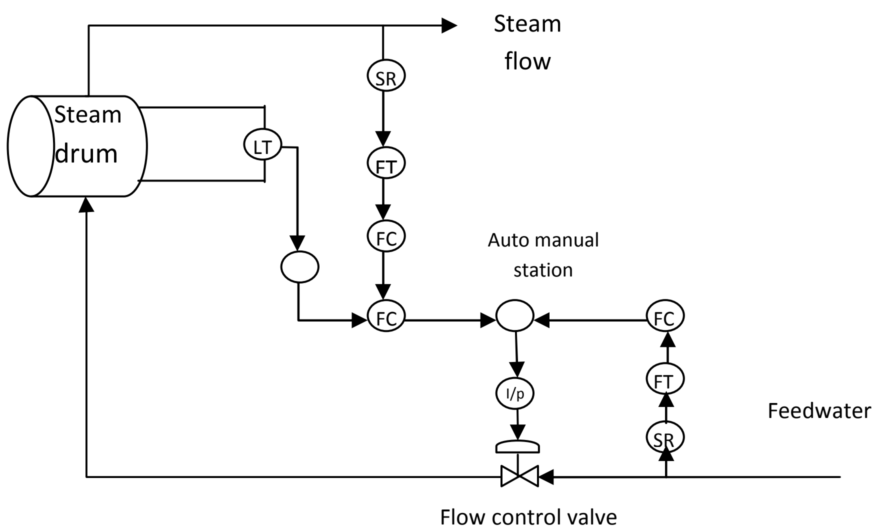

Figure: 4.12 schematic of three-element control

Chapter 5

MATLAB IMPLEMENTATION AND RESULTS

5.1 DISTILLATION COLUMN MODEL:

5.1.1 Cascade control response of Distillation Column

(The model of a distillation column is referred/taken from the text “process dynamics and control” by Dale E.Seborg, Thomas F. Edgar and Duncan A. Mellichamp [2]. Page no:558 and equation no (20-74))

Consider the wood-Berry model[13]

$$\begin{bmatrix} X_D(s) \\ X_B(s) \end{bmatrix} = \begin{bmatrix} \frac{12.8e^{-s}}{16.7s+1} & \frac{-18.9e^{-3s}}{21s+1} \\ \frac{6.6e^{-7s}}{10.9s+1} & \frac{-19.4e^{-s}}{14.4s+1} \end{bmatrix} \begin{bmatrix} R(s) \\ S(s) \end{bmatrix} + \begin{bmatrix} \frac{3.8e^{-8.1s}}{14.9s+1} \\ \frac{4.9e^{-3.4s}}{13.2s+1} \end{bmatrix} F(s)$$

The distillate and bottoms compositions (X_D and X_B) are the controlled variables; the manipulated variables are the reflux flow rate and the steam flow rate to the re boiler (R and S); and feed flow rate F is an unmeasured disturbance.

Here the Distillate composition (X_D(s)) and Reflux (R(s)) are related with the transfer function $\frac{12.8e^{-s}}{16.7s+1}$; X_D(s) and S(s) are related with the transfer function $\frac{-18.9e^{-3s}}{21s+1}$.

Similarly X_B(s) and R(s) related with $\frac{6.6e^{-7s}}{10.9s+1}$; X_B(s) and S(s) related with $\frac{-19.4e^{-s}}{14.4s+1}$.

The Cascade control loop (as shown in Figure 4.2) is implemented in matlab using simulink.

Thus the Distillation composition is controlled by manipulating the reflux flow R and the bottoms composition is controlled by the manipulating steam flow S.

In matlab simulink can be opened using the command “simulink” in the command window.

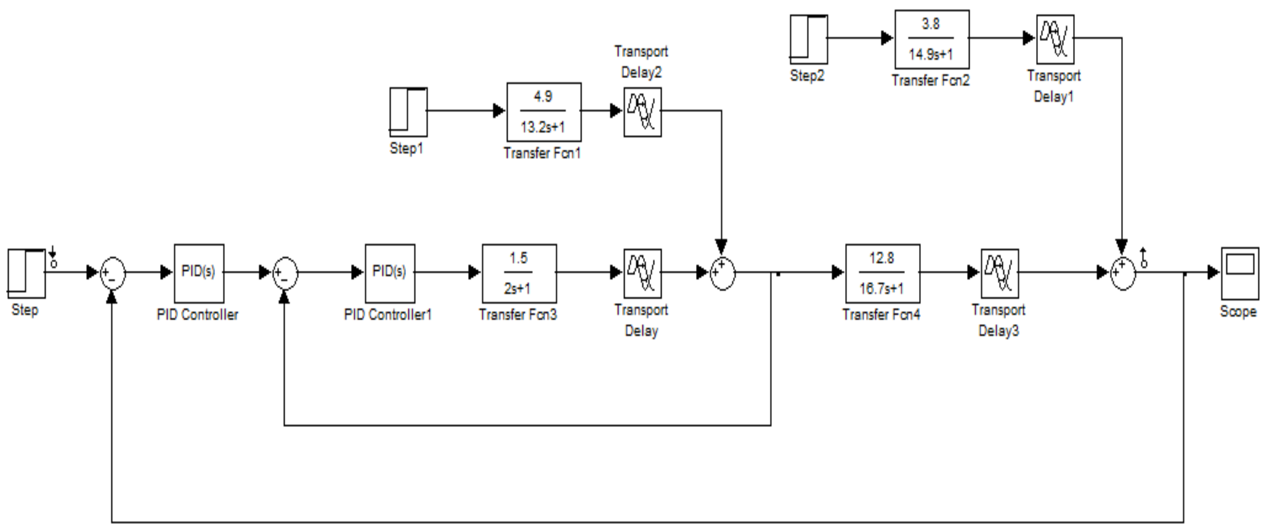

Figure:5.1 Cascade control loop using Simulink.

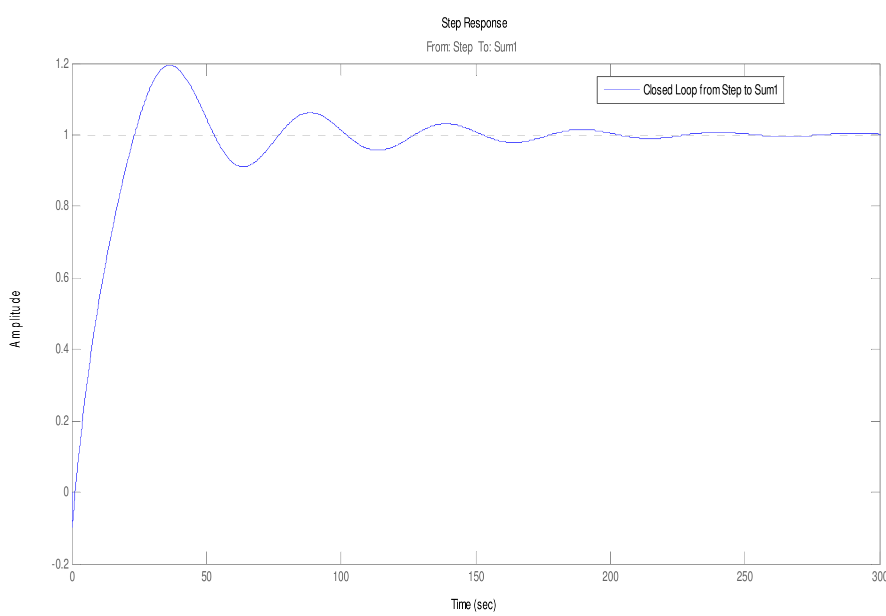

Figure:5.2 Output Response of the cascade controller.

5.1.2 *Feedforward control response of Distillation column*

Consider the wood-Berry model process and disturbance transfer functions (from the text "process dynamics and control" by Dale E. Seborg, Thomas F. Edgar and Duncan A. Mellichamp [2]. Page no: 558 and equation no (20-74))

$$G_p(s) = \frac{12.8e^{-s}}{16.7s+1} \tag{5.1}$$

$$G_d(s) = \frac{3.8e^{-8.1s}}{14.9s+1} \tag{5.2}$$

The feed-forward control system should have two transfer functions Gc(s) and Gp(s), which completes the design of feed-forward control mechanism.

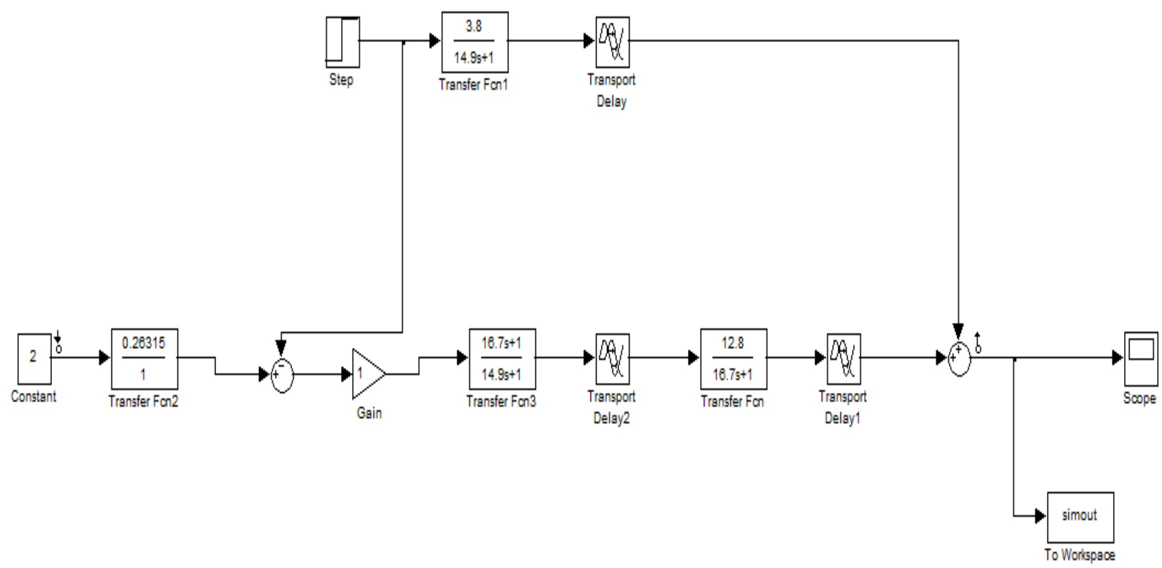

Figure:5.3 Feed-forward control loop using Simulink.

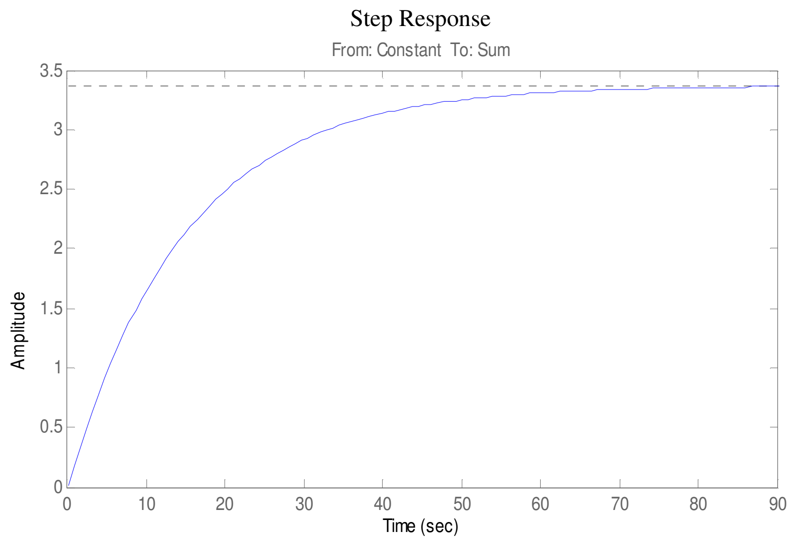

Figure: 5.4 Output Response of the feed-forward controller.

5.1.3 Comparision of output response of Cascade control and Feedforward control

I have applied cascade and feedforwad controller only Wood-Berry process model of distillation column. Step Response have been observed by simulink realisation and got the responses on the different characteristics.

We can see from the response of the cascade controller there are some oscillations, peak over shoot and large settling time present in the graph. Where as in the response of the feed-forward controller there is no overshoot came and its settling time is decreased too for the same process.

Table:1 Comparison of different parameters in Cascade and Feed forward Controllers

Sl. No	Controllers	Overshoot (%)	Settling time (min)
1.	Cascade Controller	20 %	380
2.	Feed forward Controller	0 %	85

5.2 HEAT EXCHANGER MODEL:

5.2.1 Feedback control response of Heat Exchanger

In the feedback control take control action after the output of the process, sensor senses the output and generates error signal and fed to the controller which corrects the output but the output again changed till the controlling action reaches. It higher peak over shoot that can be minimised by applying the feed-forward along with the feedback control.

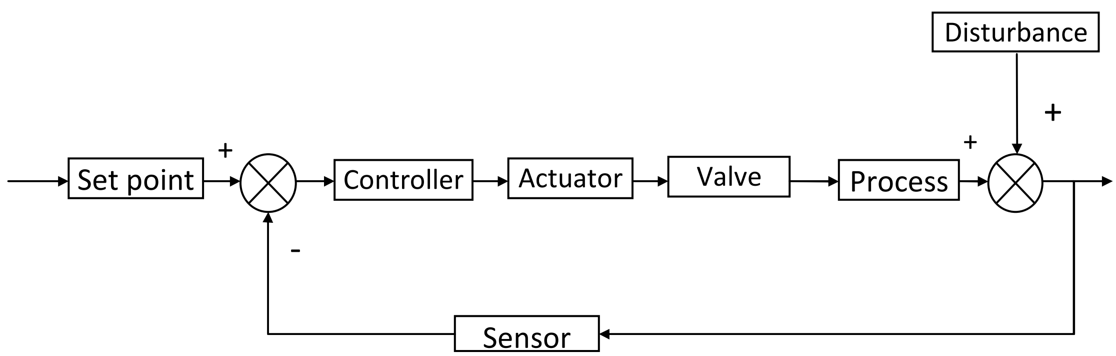

Figure: 5.5 Basic block diagram of Feedback control of heat exchanger

According to the figure and taken transfer functions

In this case the characteristic equation (l + G(s) H(s) = 0) is obtained as below.

$$900s3+420s2+43s+0.78Kc+I=0 \tag{5.3}$$

Routh Hurwitz stability criterion gives K_C as 24.45.

Auxiliary equation is given by

$$420S^2+O.798K_c+l=0 \tag{5.4}$$

After putting the value of S=jw and K_C = 24.45 we got:

w=0.218 and T=28.57

Now, K_p =14.67, τ_d =3.571, τ_i =14.28

$$G_{pid} = K_c + \frac{K_c}{\tau_i s} + \tau_d K_c S \tag{5.5}$$

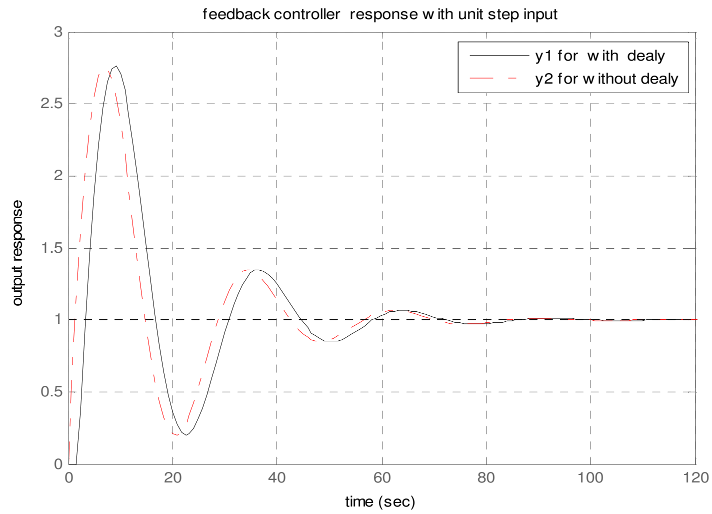

Figure: 5.6 Output Response of the feedback controller.

5.2.2 Feedback-Feedforward control response of Heat Exchanger

The main advantage of the feed-forward controller is that it analyzes the error and before disturbance will affect the process it changes the manipulating variable. The combination of feed forward and feedback reduces the over shoot.

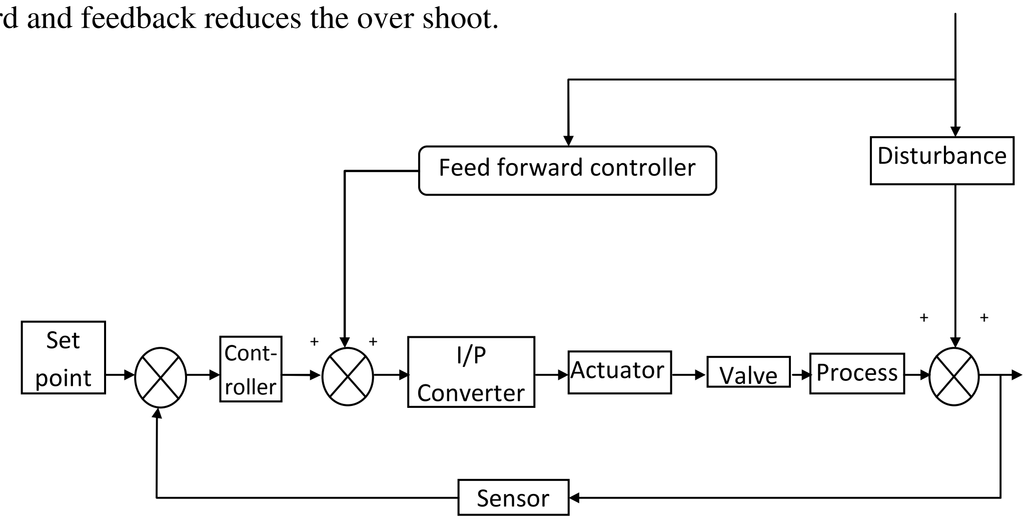

Figure: 5.7 generalized form of feedback-feed forward control

Using the equation given below we can find out the output response of the heat exchanger.

$$\bar{y} = \frac{G_p G_f (G_{c1} + G_{c2} G_{sp})}{1 + G_p G_f G_{c1} G_{m1}} \bar{y}_{sp} + \frac{G_d - G_p G_f G_{c2} G_{m2}}{1 + G_p G_f G_{c1} G_{m1}} \bar{d} \quad (5.6)$$

Where:

$$G_{c1} = \frac{87.40S^2 + 24.45S + 1.712}{S} \quad (5.7)$$

$$G_{c2} = \frac{-18.461S^2 - 6.77S - 0.205}{(30S+1)(\lambda S+1)} \quad (5.8)$$

$$G_f = \frac{0.75 \times 0.13}{(3S+1)} \quad (5.9)$$

$$G_p = \frac{50e^{-sT_d}}{(30S+1)} \quad (5.10)$$

$$G_{sp=1} \quad (5.11)$$

$$G_{m1} = \frac{0.16}{10S+1} \quad (5.12)$$

$$G_{m2} = 1 \quad (5.13)$$

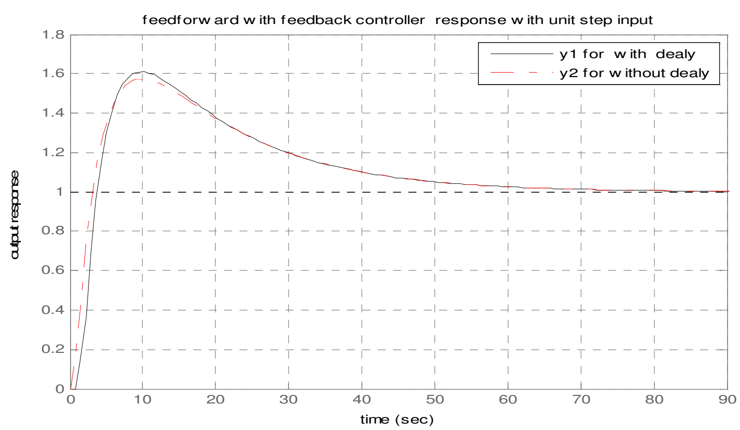

Figure: 5.8 Output Response of the feedback-feedforward controller.

5.2.3 Cascade control response of Heat Exchanger

Cascade control is used to eliminate effects of disturbance from the control-flow. The main process variable is controlled by outer loop by changing the set point for flow to an inner loop. Actual flow rate is controlled and measured by the inner loop and correct it instantly for deviation from the set-point.

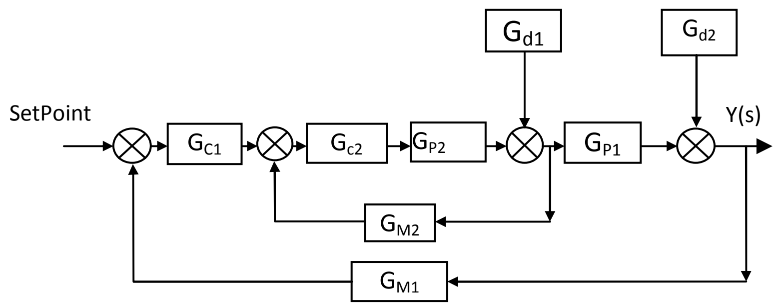

Figure: 5.9 Block diagram of cascade control of heat exchanger

Here we can find the output response using the equation no. (4.26)

$$\left(\frac{C_1}{R_1}\right) = \frac{G_{c1}G_{c2}G_{p1}G_{p2}}{1 + G_{p2}G_{m2}G_{c2} + G_{p1}G_{p2}G_{c1}G_{c2}G_{m1}}$$

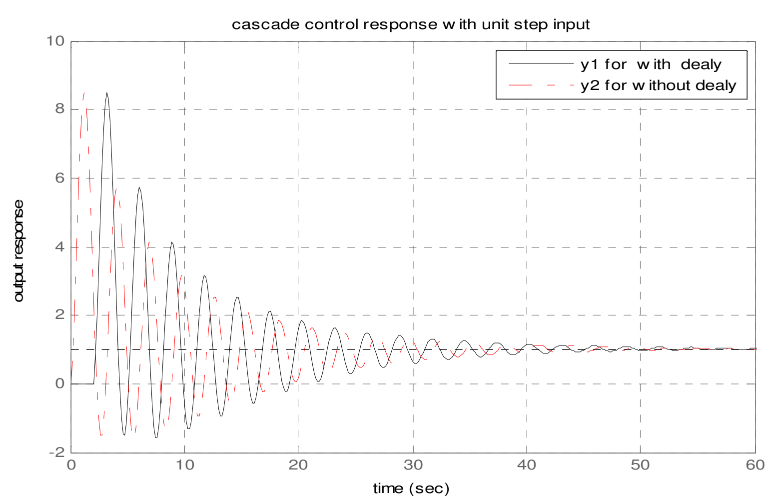

Figure: 5.10 Output Response of the cascade controller.

5.2.4 *Internal Model Control response of Heat Exchanger*

The outlet water temperature reacts faster to a decrease in the valve position (cooling) than to an increase in the steam valve position (heating). Also, according to the change in inlet flow rate, there is also changes occurs in the physical parameters such as dead time and process gain. These characteristics make the heat-exchanger to constitute performance analysis with IMC. In the IMC controller the tuning parameter is LAMBDA and I have taken the graph of the two lambda value.

Transfer function of Heat Exchanger

$$G_p(s) = \frac{50e^{-ST_d}}{(30S+1)} \tag{5.14}$$

Breaking into invertible and non invertible part

$$G_p(s) = G_p^-(s) \times \ G_p^+(s)$$

$$G_p(s) = \frac{50}{(30S+1)} \times e^{-ST_d}$$

IMC controller

$$Q(s)(IMC) = \frac{1}{G_p^+(s)} = \frac{(30S+1)}{50} \tag{5.15}$$

Making it proper by padding filter

$$Q(s) = \left(\frac{30S+1}{50}\right) \times \frac{1}{(\lambda S+1)}$$

$$Q(s) = \frac{(30S+1)}{50} \times f(s)$$

$$f(s) = \frac{1}{(\lambda S+1)}\ ,\frac{(\Upsilon S+1)}{(\lambda S+1)^2}$$

$$Q(s) = \frac{(30S+1)}{50} \times \frac{(\Upsilon S+1)}{(\lambda S+1)^2}$$

Where: $$\Upsilon = \frac{2\lambda\tau_p - \lambda^2}{\tau_p} = \frac{60\lambda - \lambda^2}{30}$$

$$\tau_d = 1\,min$$

$$Y(s) = \frac{e^{-s}}{(\lambda S + 1)^2} \times \frac{(60\lambda - \lambda^2)}{30}$$

Final output response can be obtained from this equation:

$$Y(s) = \frac{(60\lambda - \lambda^2)e^{-s}}{30(\lambda^2 + 2\lambda S + 1)} \tag{5.16}$$

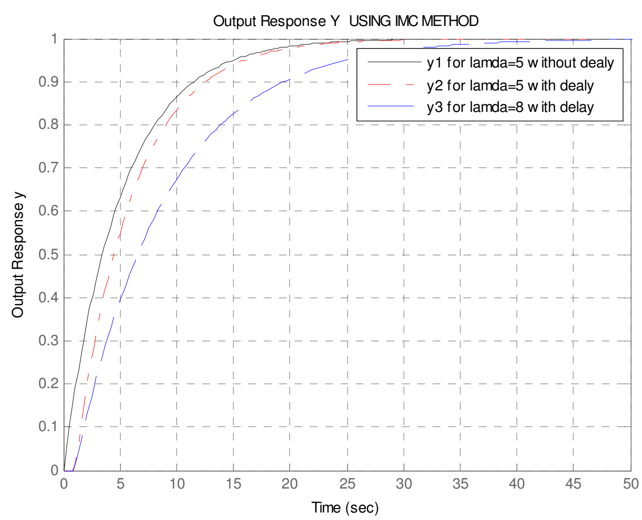

Figure: 5.11 Output Response of the IMC controller.

5.2.5 Comparision of all controller responses

After applying all the controllers on the heat exchanger system we got the different response on the following aspects such as peak over shoot, settling time, offset error etc.

1. Feedback controller: According to the Figure: (5.5) response has a large overshoot, large settling time and also response is not coming offset free.

2. Feedback-feed forward controller: According to the Figure: (5.6) due to the additional application of the feed forward controller in the forward path it improves some parameters. In comparison with the feedback controller it gives less overshoot, less number of oscillations, less offset error and good settling time mean. It gives the better set point tracking.
3. Cascade controller: According to the figure: (5.7) In this two loops are cascade and in the response there is less settling time in comparison with the feedback and feedback-feed forward controllers but the large number of oscillation is present also it is not giving the offset free output. Because of large initial fluctuation it has poor response than the feedback and feedback-feed forward controllers.
4. Internal Model Controller: According to the Figure: (5.8) the response of the Internal Model Controller has the best among all the controllers in all aspects. It has the very good set point tracking, no oscillations, no offset error, very less settling time, good response time and no overshoots in comparison with the other three controllers. It is more accurate than the other and easy to implement in Matlab.

Table: 2 Comparison of different parameters in Controllers

Sl No.	Controllers	Overshoot (%)	Settling Time (min)
1.	Feedback controller	175 %	98
2.	Feedback-feed forward Controller	60%	74
3.	Cascade Controller	**800%**	63
4.	IMC Controller	NO	28

5.3 BOILER DRUM MODEL:

5.3.1 Feedback-feed forward control response of Boiler Drum

Transfer function of the Boiler Drum process:[12]

$$G_p(s) = \frac{0.25(-S+1)}{S(0.15S+1)} \tag{5.17}$$

$$G_v(s) = \frac{1}{0.15S+1} \tag{5.18}$$

$$G_d(s) = \frac{-0.25(-S+1)}{S(2S+1)} \tag{5.19}$$

$$G_{ff} = -\frac{G_d(s)}{G_p(s)} = 1 \tag{5.20}$$

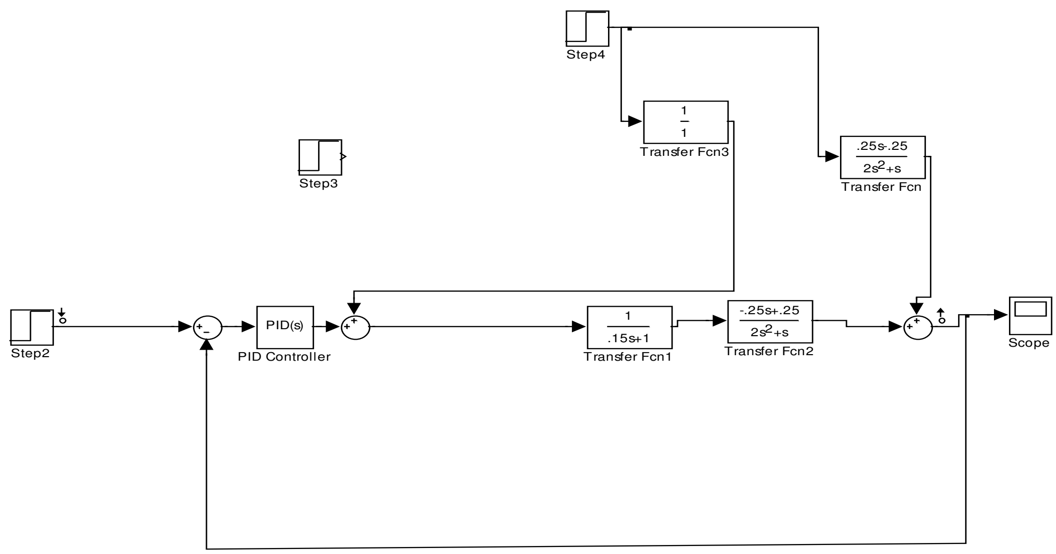

Figure: 5.12 Feedback-feed forward simulink implementation

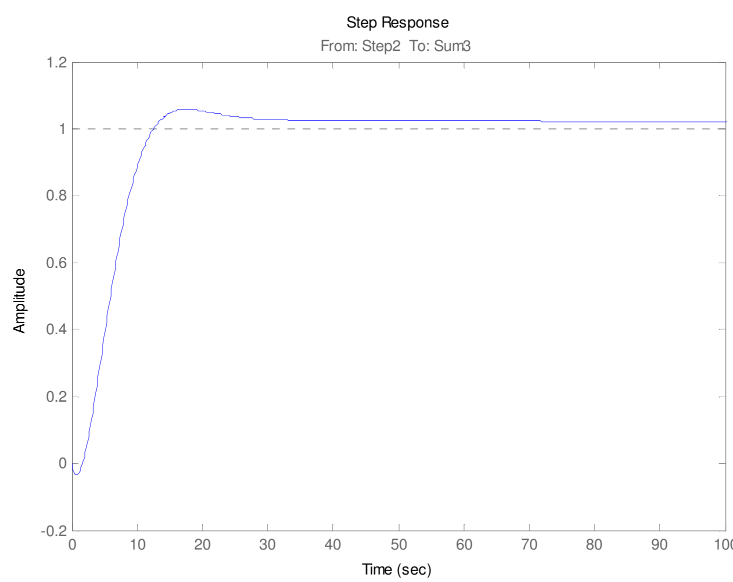

Figure: 5.13 Output response of Feedback-feed forward control of Boiler Drum

5.3.2 Combination of Feedback-feed forward and cascade response of Boiler Drum

Same transfer function of Boiler Drum is used in outer loop and transfer function of valve is taken in the inner loop. These transfer functions are shown in equation number (5.15, 5.16). Gain=1 is taken as a secondary controller.

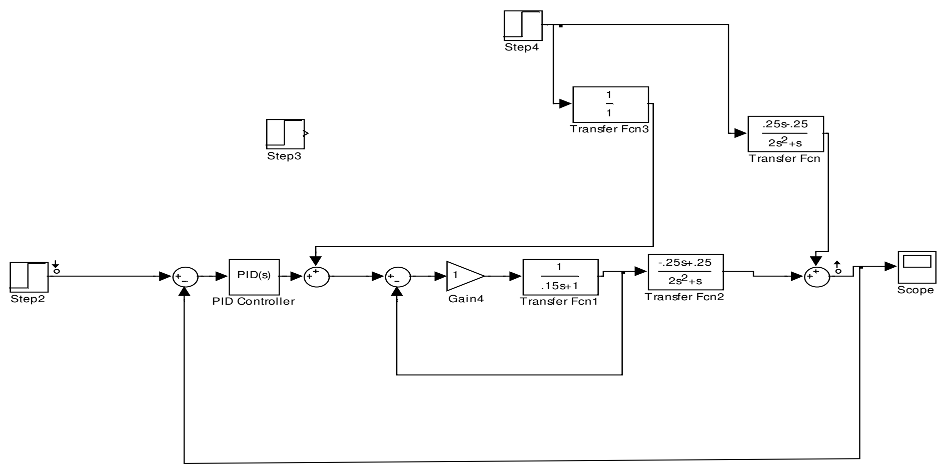

Figure: 5.14 Feedback-feed forward and cascade simulink implementation

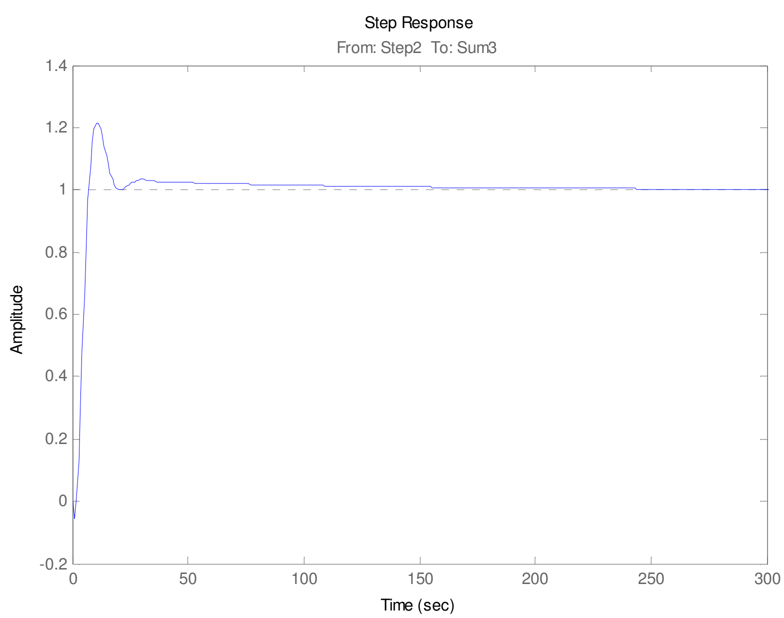

Figure: 5.15Output response of Feedback-feed forward and cascade control of Boiler Drum

5.3.3 Internal Model Control response of Boiler Drum

I have used the transfer function of the boiler drum from the [12]. They have done the control on feed water flow to water level. And I used this transfer function to apply IMC control.

Transfer function of the Boiler Drum:

$$G_p(s) = \frac{0.25(1-S)}{S(2S+1)} \tag{5.21}$$

After applying the same procedure as applied in the IMC control of the heat exchanger system I got the equation of output:

$$Y(s) = \frac{(1-S)}{(\lambda S+1)^2} \tag{5.22}$$

I have got the two responses, controller output response and output response of Boiler Drum. In the plot of Controller output there are three composite curves with different lambda value. It shows the better response as the value of lambda decreases and manipulated variable saturates early.

And in the plot of output response there are also three composite curves with different lambda value. In this response gives the better response as the value of lambda increased means curve approaches the set point very early.

<u>Table: 3 Comparison of different parameters in Controllers</u>

Sl No.	Controllers	Peak overshoot(%)	Settling time(min)
1.	Feedback-feedforward	10%	43
2.	Feedback-feed forward-Cascade	20%	30
3.	IMC	NO	17

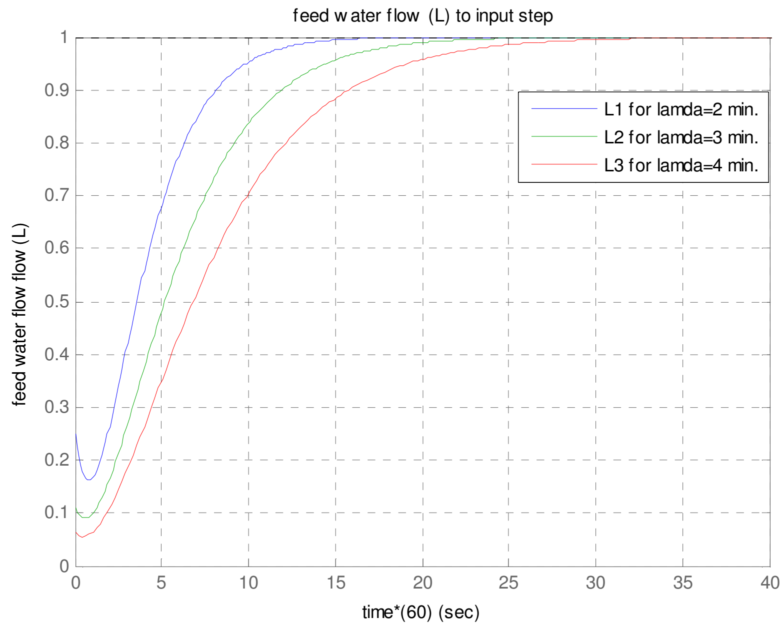

Figure: 5.16 Controller output response response of Boiler Drum

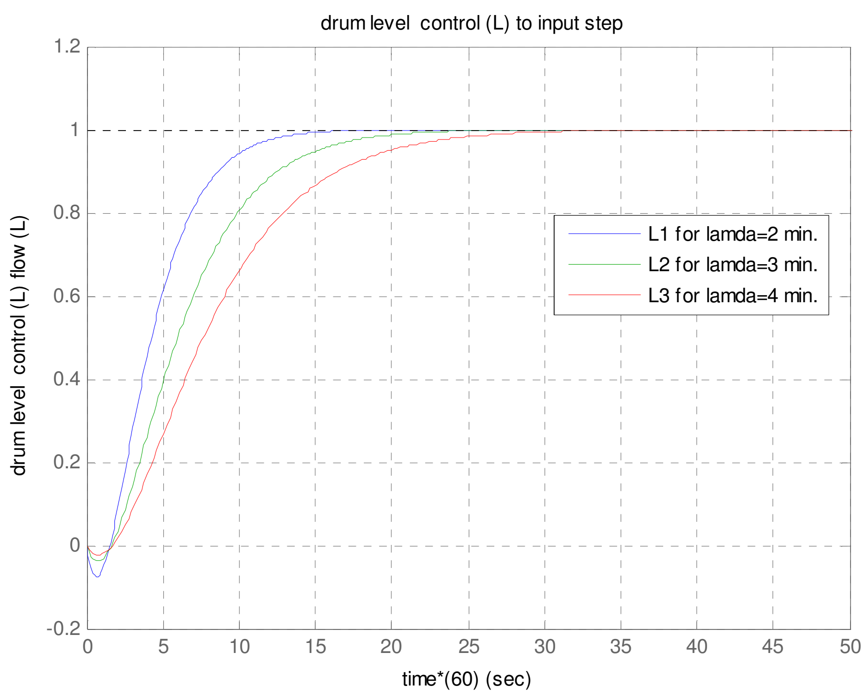

Figure:5.17 Output response of the Boiler Drum

Chapter 6

CONCLUSION AND SCOPE FOR FUTURE

6.1 CONCLUSION

I have used the controllers Feedback, Feed forward, Combination of Feedback and Feed forward, Cascade and Internal Model Controllers to see the responses of the controllers. On the Distillation column I applied cascade and feed forward , on heat exchanger I applied Feed forward, Combination of Feedback and Feed forward, Cascade and Internal Model Controllers and on Boiler Drum only IMC.

In distillation column result there is large settling time so the use of feed forward controller is better because its settling time as well as its overshoot is also improved.

In Heat Exchanger system I observed that among all the controllers response of the IMC controller is best. In the response of cascade control on heat exchanger there are so much initial fluctuation occurs but the settling time is better the feedback and the combination of feedback and feed forward.

In Boiler Drum three controllers have applied and got the output response of that among that there is no overshoot in IMC and in feedback with feed forward 10%. In the combination of feedback, feed forward and cascade the over shoot is increased. In the settling time IMC is better than other two.

6.2 FUTURE SCOPE

The following are some of the prospects for future work:

1. Further implementation IMC based cascade controller and IMC based feerforward controller.
2. The feedback and feed forward controllers can be used for Multivariable systems. It can be further implemented for the robust control system using IMC.

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DISSEMATION

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